STUDY RELATING TO THE TECHNICAL AND ECONOMIC FEASIBILITY OF LIGNITE LIQUEFACTION IN SASKATCHEWAN

Volume I

SUMMARY AND PROCESS COMPARISON

Final Report
by

TECHNICAL DEVELOPMENT DEPARTMENT, SNC Inc.

for

SASKATCHEWAN OIL AND GAS CORPORATION

and

CANMET, ENERGY, MINES AND RESOURCES, CANADA

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J.A. Mikhlin, Eng, Ph.D.

EXECUTIVE SUMMARY

This study is carried out by Coal Conversion Group, Technical Development Department, SNC Inc., on behalf of Saskatchewan Oil and Gas Corporation (Saskoil) and Canada Centre for Mineral and Energy Technology (CANMET).

Review and Evaluation of Advanced Processes

The study comprises review and evaluation of advanced direct and indirect coal liquefaction and pyrolysis processes considered for commercialization by 1990 among those currently being developed in the U.S.A., Federal Republic of Germany, and the U.K. Also, a commercially proven SASOL-type Fischer-Tropsch process (South Africa) is considered. The ultimate objective is to identify the most promising coal liquefaction process(es) for the designated Saskatchewan lignite. The processes evaluated are:

- Direct Coal Liquefaction
  - H-Coal
  - Ruhrkohle/Veba Oel (RAG/Veba)
  - Rheinische Braunkohlenwerke (Rheinbraun)
  - Exxon Donor Solvent (EDS)
  - Liquid Solvent Extraction (LSE)
- Indirect Coal Liquefaction
  - Coal-based plant for the production of pure methanol
  - Coal-based Methanol-to-Gasoline plant (Mobil MTG)
  - SASOL-type plant with Fischer-Tropsch Synthesis
- Coal Pyrolysis
  - Lurgi/Ruhrgas (L-R) process

Review and Evaluation of Advanced Processes (cont'd)

The evaluation was carried out in two stages. Stage I covers technical assessment, prescreening and ranking of the processes in each category for the designated Saskatchewan lignites, i.e. from the Estevan and the Willow Bunch deposits. Stage II covers techno-economic assessment of processes selected in Stage I but for Willow Bunch lignite only.

The characteristic features of designated lignites are high ash content (15.4 wt% MF for Estevan and 21.0 wt% MF for Willow Bunch lignite) and high moisture content (35.0 wt% for both lignites).

Indirect coal liquefaction requires initial gasification of coal to synthesis gas and then converting the synthesis gas produced to desired liquid product. Therefore, the evaluation of any indirect coal liquefaction route will require, first of all, a selection of a proper gasifier for a designated coal. The following advanced gasifications configurations are considered for evaluation and prescreening in Stage I.

- Fixed-Bed Type
  - Lurgi (dry bottom)
  - British Gas/Lurgi (slagging)
- Entrained-Flow Type
  - Koppers-Totzek
  - Texaco
- Fluidized-Bed Type
  - High-Temperature Winkler (HTW)
  - Westinghouse

EXECUTIVE SUMMARY (cont'd)

Process Prescreening of Advanced Process (Stage I of the Study)

The technical assessment in Stage I highlights the specific features of each process and includes a review of process history and development, relevant experimental results pertaining to processing of lignites. An overall configuration is presented for a conceptual, commercial plant processing 25,000 tonnes per day of as-received lignite. Product yield and distribution, overall and desired product efficiencies for a self-sufficient integrated plant, and the quality of the products are also investigated. Current status of each process, areas of technical concerns and commercialization constraints are determined. The product criteria for Stage I of the study are as follows:

- synthetic crude (naphtha, middle oil, heavy oil) via direct coal liquefaction and pyrolysis processes
- transportation fuel (gasoline, diesel, jet fuel) via indirect coal liquefaction but with emphasis on gasoline production
- pure methanol (99.5% purity) as transportation fuel and blendstock for gasoline via indirect coal liquefaction.

In direct coal liquefaction, the conversion of coal to desired liquid process inevitably involves handling, separation, and disposal of ash. The high-ash content of Saskatchewan lignites have a negative effect on the overall product yield and efficiency of the process. This is particularly significant for one-step coal liquefaction processes where non-mechanical solid/liquid separaton methods (flash separation, vacuum distillation) are used.

EXECUTIVE SUMMARY (cont'd)

Process Prescreening of Advanced Process (Stage I of the Study) (cont'd)

One of the ways to reduce the negative effect of ash is to use a two-step approach where coal is first dissolved in a coal-derived solvent (extraction stage) and the extract is separated from the mineral matter by mechanical separation such as filtration, centrifugation, etc. The ash-free extract is then selectively hydrocracked (hydrogenation stage) in a catalytic reactor.

Among the processes considered, the two-step LSE process using filtration for solid/liquid separation, appears to have highest priority and was selected for further techno-economic assessment of Willow Bunch lignite (Stage II of the Study), although its commercialization prospects are somewhat uncertain. The next best candidates are the H-Coal and the EDS processes.

If the ash content can be economically reduced by beneficiation/pretreatment to about 10 wt% MF coal, the one-step approach may be attractive. Also, new developments in H-Coal (two-step approach) and EDS (bottoms recycle mode) processes should also be considered in future work, since complete data on these development are not yet available.

For indirect coal liquefaction, entrained-flow Koppers-Totzek gasifier appears to have the highest priority, at present, for production of syngas from Saskatchewan lignite since the methane content in the gas produced is negligible and there are no restriction on handling of coal fines. This gasifier is commercially proven on lignites with ash content upto 32 wt% MF coal.

Process Prescreening of Advanced Process (Stage I of the Study) (cont'd)

The second generation fluidized bed gasifiers, High-Temperature Winkler (HTW) and Westinghouse, appears to be next best for gasification of Saskatchewan lignites. Experimental tests required by process developer to confirm the gasification behaviour of Saskatchewan lignites are beyond the scope of the present study. Based on the experimental results, a techno-economic comparison of these gasifiers configuration with entrained-flow K-T gasifier is recommended for future work.

In pyrolysis route, the Lurgi-Ruhrgas (L-R) process may be viable for Estevan lignite if a limited amount of liquid product is acceptable and surplus char can be marketed or utilized effectively in a power plant. For Willow Bunch lignite, the L-R process is not viable because of low predicted liquid product yields based on laboratory pyrolysis tests (Fischer-Schrader Assay). The L-R process was not considered for techno-economic assessment (Stage II of the Study).

Techno-Economic Assessment of Preselected Process (Stage II of the Study)

In the Stage II, techno-economic assessment was carried out for preselected processes, i.e., LSE in the direct coal liquefaction route and three process alternatives in the indirect coal liquefaction (methanol plant, Mobil MTG plant and SASOL-type plant) with Koppers-Totzek gasifier for production of synthesis gas. For consistent comparison of direct and indirect coal liquefaction routes, the transportation fuels such as gasoline, diesel and jet fuel with emphasis on the production of gasoline are considered for all cases (except methanol plant). The following financial assumption and economic criterias are considered for estimating the capital, operating, maintenance, and the product costs.

EXECUTIVE SUMMARY (cont'd)

Techno-Economic Assessment of Preselected Process (Stage II of the Study) (cont'd)

- Coal feed: 25,000 t/d of as-received Willow Bunch lignite
- Plant location: within 16 km (10 miles) of the coal mine
- Constant Canadian dollars, mid-1982
- 4-year construction period
- Plant life: 30 years
- Straight-line depreciation with zero salvage value
- Cost of capital financing: 10% per annum
- Cost of coal: \$9.75/tonne (includes \$0.28/tonne transportation cost)
- Plant service factor: 0.85 (310 operating days per year).

Product costs for each of the four preselected processes are calculated on an "Energy Value" and "Product Value" bases. The first method is based on heating value of the product only and is simply the unit energy cost (\$/10^6 kJ). The second method also considers the quality and market value of the product. The latter approach was developed by the Engineering Society Commission on Energy Inc. (ESCOE).

In calculating the product cost, credits are taken for byproducts such as sulfur at \$65/t, ammonia at \$300/t and crude-mixed phenols at \$250/t.

EXECUTIVE SUMMARY (cont'd)

Techno-Economic Assessment of Preselected Process (Stage II of the Study) (cont'd)

A summary of techno-economic results is shown in Table 1. For LSE process, two cases representing the lower (65% on MAF basis) and upper extraction levels (82% MAF basis) attainable in the reactor are presented.

Analyzing the product output and distribution, it may be seen that different products in varying amounts are produced in the processes. However, all the products meet the standard specifications. Comparison of total amount of desired liquid products (gasoline, jet fuel, diesel and fuel oil) show that the LSE process has the highest yield (4154 to 4387 t/d) followed by Mobil-MTG (2245 t/d) and finally the lowest yeild for SASOL-type F-T plant (1332 t/d). However, all the liquid produced in Mobil-MTG is gasoline while in the LSE process 36 wt% is gasoline, 34 wt% jet fuel and the remaining 30 wt% is diesel fuel. The product yield from methanol plant are not compared with other processes because of a different nature of the product. The performance data of methanol is discussed separately.

The total product yield is highest for the LSE process followed by Mobil-MTG and the lowest for SASOL-type F-T plant. Similar trends are observed in overall and liquid product efficiencies*.

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* In this study, a modified version of thermal efficiency widely used in the energy-related process is employed. The modified defination of the efficiency is the total gross-calorific-value (GCV) of the product divided by the total input energy of the lignite feed and electric power.

Table 1 Summary of Techno-Economic Assessment

	LSE Process		Mobil MTG Plant	SASOL-Type F-T Plant	Methanol Plant
	65% Extraction MAF Basis	82% Extraction MAF Basis			
Product Output and Distribution, t/d					
Fuel Gas	-	-	64	784	-
Propane	-	-	117	69	-
Propane/Butane (LPG)	389	410	-	-	-
Butane	-	-	186	11	-
Heavy Fuel Oil	-	-	-	63	-
Jet Fuel	1,402	1,481	-	-	-
Diesel	1,262	1,334	-	214	-
Gasoline	1,490	1,572	2,245	1,055	-
Mixed Alcohols	-	-	-	133	-
Methanol	-	-	-	-	5,714
Desired Liquid Product Yield*					
bbl/t as-received coal to plant	1.11	1.17	0.77	0.46	1.75
bbl/t MF coal to plant	1.71	1.80	1.19	0.71	2.69
bbl/t MAF coal to plant	2.16	2.27	1.50	0.91	3.41
Total Product Yield**					
bbl/t as-received coal to plant	1.12	1.19	0.94	0.94	1.75
bbl/t MF coal to plant	1.73	1.83	1.45	1.44	2.69
bbl/t MAF coal to plant	2.19	2.32	1.84	1.82	3.41
Product Efficiency					
Desired Liquid Product	51.0	54.0	31.4	18.0	-
Total Liquid Product	51.0	54.0	31.4	18.9	39.1
Overall Product	57.5	60.7	43.5	36.0	39.5

Table 1 Summary of Techno-Economic Assessment (cont'd)

		LSE Process		Mobil MTG Plant	SASOL-Type F-T Plant	Methanol Plant
		65% Extraction MAF Basis	82% Extraction MAF Basis			
Capital Investment, $ million		2,639	2,639	2,213	2,545	1,926
Investment, $/daily bbl of desired liquid product*		95,054	90,000	114,750	219,245	44,010
Investment, $/daily bbl of total products**		93,740	88,765	93,770	109,035	44,010
Average Product Cost, $\$/10^6$kJ						
Energy Value Basis		9.48	8.97	11.28	15.27	11.14
Product Value Basis		12.43	11.76	13.92	22.72	8.70
Individual Product Cost (Product Value Basis)						
Fuel Gas	$\$/10^6$kJ	-	-	4.87	7.95	-
Propane	$/bbl	-	-	22.37	36.61	-
Propane/Butane	$/bbl	22.03	20.84	-	-	-
Butane	$/bbl	-	-	35.64	62.45	-
Heavy Fuel Oil	¢/L	-	-	-	83.54	-
Jet Fuel	¢/L	29.3	27.7	-	-	-
Diesel	¢/L	41.1	38.9	-	66.1	-
Gasoline	¢/L	43.6	41.2	45.3	72.7	-
Mixed Alcohols	¢/bbl	-	-	-	86.4	-
Methanol	$/bbl	-	-	-	-	32.97
Components of Product Cost, %						
Coal Cost		13.8	13.8	15.1	13.4	16.9
Operating and Maintenance Cost		32.6	32.6	34.1	34.9	33.8
Capital Cost		53.6	53.6	50.8	51.7	49.3

* Desired liquid product are gasoline, diesel, jet fuel (Pure methanol in methanol plant)
**Total product refers to all the product listed above.

EXECUTIVE SUMMARY (cont'd)

Techno-Economic Assessment of Preselected Process (Stage II of the Study) (cont'd)

The total capital investment for a self-sufficient, integrated plant processing 25,000 t/d of as-received Willow Bunch lignite is nearly the same for LSE and SASOL-type F-T plants and is lowest for Mobil-MTG plant. However, investment per daily production of desired products is lowest for LSE, followed by Mobil-MTG and F-T plants. In other words, a plant based on equal output of products would have the lowest capital investment for the LSE process and highest capital investment for F-T plant. If gasoline was the only desired product, the capital investment per daily production of gasoline would be by far the lowest for Mobil-MTG plant.

The product cost for LSE and Mobil-MTG process are nearly the same but for SASOL-type plant it is quite high. However, the breakdown of product cost into main components show that they are of same proportion in all cases.

The LSE process has been demonstrated on a continuous bench-scale unit having a capacity of processing 0.7 t/d of coal. Some of the processing steps have also been separately demonstrated on a larger scale such as 2 t/d solvent extraction unit. To demonstrate the process on a larger scale, a decision was made to build a pilot plant of 22.7 t/d (25 short ton/day) of coal. The current economic difficulties coupled with temporary world oil glut has delayed the development of LSE process, as well as development of other processes, including several the processes. The Mobil-MTG process still has better commercialization prospects since the only part of this process that has not been demonstrated commercially is the methanol to gasoline conversion. On the other hand, the SASOL-type F-T plants are commercially proven.

Techno-Economic Assessment of Preselected Process (Stage II of the Study) (cont'd)

It is concluded that LSE and Mobil-MTG are comparable. The SASOL-type plant, although commercially proven, has very low desired product yield and is expensive. To further distinguish between LSE and Mobil-MTG process, a detailed conceptual design of a commercial plant and economic and financial assessment (including sensitivity analysis) with a market survey for transportation fuels in the Province of Saskatchewan is recommended.

For the methanol plant, the product yield is 2.69 bbl/t MF coal whereas the product efficiency is relatively low (about 39%), primarily due to the indirect method of liquefaction. Capital cost are relatively lower and the product costs are comparable to present-day market prices (which are based on use of methanol as chemical feedstock only). The disadvantages are that methanol has low energy density and there are several practical problems associated with its use as a fuel, such as uncertain market demand, distribution and storage difficulties, cold starting problems, lower tolerance for water, etc.

Preliminary Evaluation of New Processes or Concepts

In addition to assessment of advanced direct and indirect coal liquefaction and pyrolysis processes (which is the main objective of the study), a preliminary review of new processes or concepts considered for commercialization around 1995 was carried out for the designated Saskatchewan lignites. Seven direct coal liquefaction processes (Dow, CO-Steam, CSF, TSL, SRC, Z-H, SRL), three coal/oil coprocessing options (Coil, JSCL, Veba), one pyrolysis process (Occidental), and two gasifiers (Shell, U-Gas) for indirect coal liquefaction were evaluated.

EXECUTIVE SUMMARY (cont'd)

Preliminary Evaluation of New Processes or Concepts (cont'd)

The main objectives of this review are to determine the current state-of-the-art of each process, their current demonstration scale, and R&D status (including funding status).

The results of the review are that some of the processing steps/features are attractive for further development in coal liquefaction, including Saskatchewan lignites. However, in most cases, process development work has been abandoned due to financial constraints. Other processes have been funded till the end of 1982. In view of the slowdown and present oil glut in the world, the prospects of these processes for commercialization around 1995 seems uncertain.

Co-processing option has potential merits, especially for the Province of Saskatchewan where lignite and heavy oil are available. However, at present, the development work is on the autoclave/bench-scale level. Further R & D work is required.

Pyrolysis and Hydroliquefaction Tests

An independent experimental program was also carried out concurrently with this study by the Chemical Technology Division of B.C. Research and the results are presented in a separate report entitled, "Liquefaction and Pyrolysis Tests on Samples of Estevan and Coronach Lignites", Project Report 2-02-691, March 1982. An Australian Morwell lignite was also tested for comparative purposes.

Modified Fischer-Schrader Assay tests for pyrolysis indicate low tar yields for Willow Bunch (Coronach) lignite (≏ 6-7 wt% MAF coal) and somewhat higher tar yields for Estevan lignite (≏ 9.5 wt% MAF coal). Char yields for both lignites are high (64-66 wt% maf coal).

Pyrolysis and Hydroliquefaction Tests (cont'd)

Direct hydroliquefaction tests using a tubing bomb experimental unit indicate wide variations in product yields depending on experimental conditions. However, in terms of total liquid yields the lignites were ranked by B.C. Research as Morwell > Estevan ≥ Willow Bunch. In terms of distillate yeilds (< 525°C), the lignites are nearly the same.

Study Report Organization

The complete report on the study comprises five volumes:

Volume I - Summary and Process Comparison

Volume II - Assessment of Advanced Direct Coal Liquefaction and Pyrolysis Processes Considered for Commercialization by 1990

Volume III - Assessment of Advanced Indirect Coal Liquefaction Processes Considered for Commercialization by 1990

Volume IV - Review of Coal Liquefaction Technologies and Concepts considered for Commercialization around 1995

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1.0 INTRODUCTION

1.0 INTRODUCTION

This study was carried out by the Coal Conversion Group, Technical Development Department of SNC Inc. to evaluate various coal liquefaction processing alternatives for two designated Saskatchewan lignites, i.e., from the Estevan and Willow Bunch deposits. The processes evaluated were subdivided into two categories.

- processes considered for commercialization by 1990, and
- processes considered for commercialization around 1995.

The scope of the work for the advanced processes (commercialization by 1990) included review and evaluation of direct and indirect coal liquefaction and pyrolysis processes listed in Table 1-1. The evaluation was made in two stages. Stage I covers technical assessment, prescreening and ranking of the processes for both Estevan and Willow Bunch lignites. Stage II covers techno-economic assessment of processes selected in Stage I but for Willow Bunch lignite only. The ultimate objective is to determine a promising process(es) for Saskatchewan lignites.

The scope of the work in the second category (commercialization around 1995) is to ascertain state-of-the-art technology and review update of the processes and concepts listed in Table 1-2. A summary of the results is presented in Chapter 7 of this volume, and a full description of the evaluation is given in Volume IV.

In addition, an independent experimental program was carried out concurrently with this study to determine comparative reactivity of Estevan and Willow Bunch (Coronach) lignite. These tests were

Table 1-1 Processes Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization by 1990

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
H-Coal	Hydrocarbon Research Inc.	U.S.A.
RAG/Veba	Ruhrkohle AG and Veba Oel AG	F.R.G.
Rheinbraun	Rheinische Braunkohlenwerke AG	F.R.G.
Exxon Donor Solvent (EDS)	Exxon Research and Engineering Company	U.S.A.
Liquid Solvent Extraction (LSE)	National Coal Board, Coal Research Establishment	U.K.
	COAL PYROLYSIS	
Lurgi-Ruhrgas	Lurgi Mineralöltechnik GmbH and Ruhrgas AG	F.R.G.
	INDIRECT COAL LIQUEFACTION	
SASOL-Type Fischer-Tropsch	SASOL Ltd.	South Africa
Mobil Methanol-to-Gasoline (M-MTG)	Mobil Research and Development Corporation	U.S.A.
	Gasifiers for Production of Synthesis Gas	
Lurgi (dry bottom)	Lurgi Mineralöltechnik AG	F.R.G.
British Gas/Lurgi (slagging)	British Gas Corporation and Lurgi Mineralöltechnik AG	U.K. F.R.G.
Koppers-Totzek	Krupp-Koppers GmbH	F.R.G.
Texaco	Texaco Research and Development Corporation	U.S.A.
Winkler (high temperature)	Rheinische Braunkohlenwerke AG	F.R.G.
Westinghouse	Westinghouse Electric Corporation	U.S.A.

Table 1-2 Processes and Concepts Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization around 1995

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
Zinc Halide	Consolidation Coal Co.	U.S.A.
Dow Coal Liquefaction	Dow Chemicals Co.	U.S.A.
Consol Synthetic Fuel (CFF)	Consolidation Coal Co.	U.S.A.
Solvent Refined Coal (SRC-1)	International Coal Refining Co.	U.S.A.
Solvent Refined Lignite (SRL)	University of North Dakota	U.S.A.
CO-Steam	Grand Forks Energy Technology Center	U.S.A.
Two-Step Liquefaction	Cities Service Co. C-E Lummus Co.	U.S.A.
	COAL PYROLYSIS	
Occidental Pyrolysis	Occidental Research Corporation	U.S.A.
	INDIRECT COAL LIQUEFACTION	
	Gasifiers for Production of Synthesis Gas	
Shell Gasifier	Shell International Petroleum	The Netherlands
U-Gas	Institute of Gas Technology	U.S.A.
	CO-PROCESSING	
Coil	Hydrocarbon Research Inc.	U.S.A.
Solvolysis	Mitsubishi Heavy Industries Ltd.	Japan
Veba Coal/Oel Co-processing	Veba Oel AG	F.R.G.

1.0 INTRODUCTION (cont'd)

made by the Chemical Technology Division of B.C. Research Laboratories and the results are presented in a separate report, "Liquefaction and Pyrolysis Tests on Samples of Estevan and Coronach Lignites", Project Report No. 2-02-691, March 1982. A summary of the results is presented in Chapter 7.

1.1 OBJECTIVES OF THE STUDY

The main objectives of the study for advanced processes considered for commercialization by 1990 were identified as follows:

Stage I, Process Evaluation and Prescreening

- to determine state-of-the-art technology and predict behavior of Estevan and Willow Bunch lignite in the processes based on available data and previous experience of the process developers with similar coals.
- prescreening and ranking of the processes for further techno-economic assessment.

Stage II, Techno-Economic Assessment of Preselected Processes

- techno-economic assessment and comparison of pre-selected processes using Willow Bunch lignite as feedstock.
- selection of the promising process(es) for liquefaction of Willow Bunch lignite to produce synfuels substituting conventional transportation fuels.
- to determine constraints toward commercialization for selected process(es).

1.2 STUDY REPORT ORGANIZATION

The study report comprises five volumes:

Volume I - Summary and Process Comparison

Volume II - Assessment of Advanced Direct Coal Liquefaction and Pyrolysis Processes Considered for Commercialization by 1990

Volume III - Assessment of Advanced Indirect Coal Liquefaction Processes Considered for Commercialization by 1990

Volume IV - Review of Coal Liquefaction Technologies and Concepts Considered for Commercialization around 1995

Volume V - Process Developers' Data and Information

This Volume I, contains:

- detailed criteria for process ranking and selection (Chapter 2)
- overview of direct and indirect liquefaction and pyrolysis processes considered for commercialization by 1990, their status and suitability for Saskatchewan lignites followed by process comparison and ranking (Chapters 3,4,5 respectively).
- Comparison of the results of techno-economic assessments of the processes selected in the prescreening stage, identification of the most promising process(es) for liquefaction of Willow Bunch lignite and recommendations for future work (Chapter 6,7).

1.2 STUDY REPORT ORGANIZATION (cont'd)

A summary of the results of Volume IV, "Review of Coal Liquefaction Technologies and Concepts Considered for Commercialization around 1995", is presented in Chapter 7 of this Volume I. A summary of an independent experimental program carried out by B.C. Research Laboratories (Report No. 2-02-691, March 1982) is also included in Chapter 7.

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION

Technical and economic criteria approved by the Saskatchewan Coal Liquefaction Technical Committee (the Study Committee) for direct and indirect liquefaction and pyrolysis processes are presented below:

2.1 FEEDSTOCK SPECIFICATIONS

Two Saskatchewan lignites, from the Estevan and Willow Bunch deposits were designated as feedstocks for "prescreening" of coal liquefaction and pyrolysis processes (Stage I of the Study). Typical ultimate, proximate and petrographic analysis of these designated lignites are summarized in Table 2-1. Both coals have high moisture and ash content. While average moisture content is the same for both coals, the ash content of Willow Bunch lignite and inertinite content are higher.

For techno-economic assessment of prescreened processes (Stage II of the Study), Willow Bunch lignite was designated as the feedstock. Detailed analyses of Willow Bunch lignite showing the "best" and "worst" coal properties are given in Appendix A.

2.2 PRODUCT CRITERIA

For prescreening of the processes (Stage I of the Study) the product criteria are:

- synthetic crude (naphtha, middle oil, heavy oil) via "direct" coal liquefaction.

2.2 PRODUCT CRITERIA (cont'd)

- synthetic crude (naphtha, middle oil, heavy oil) via pyrolysis.
- gasoline as the main product via "indirect" coal liquefaction.
- pure methanol (99.5% purity) as transportation fuel or blendstock for gasoline via indirect coal liquefaction.

In order to allow a consistent comparison of the techno-economic assessments of preselected processes (Stage II of the Study), transportation fuels such as gasoline, diesel and jet fuel are considered as desired end-products for both "direct" and "indirect" coal liquefaction processes. However, emphasis is made on production of gasoline. Light hydrocarbons (C_1-C_4) are considered as co-products.

2.3 OVERALL AND LIQUID PRODUCT EFFICIENCY

A general definition of thermodynamic efficiency which may be applied to any conversion process, is in the simplest form, the total output energy divided by total input energy. Heat is taken to be the total energy content as measured by its calorific value plus any sensible heat above the datum temperature. It also includes the thermal equivalent of electricity, or pressure energy, or any other energy form that is imported or exported across the process boundaries.

In this study, a modified version of efficiency widely used for energy-related processes has been considered. The modified definition of "overall" product efficiency is the total gross-calorific-value (GCV) of all products divided by the total energy input of coal feed and electric power. For processes,

Table 2-1 Typical Ultimate, Proximate, and Petrographic Analysis of Saskatchewan Lignites

	Estevan	Willow Bunch
Moisture (as received), wt%	35.0	35.0
Gross Calorific Value (GCV) kJ/kg MF coal	24,270	20,380
Proximate Analysis, wt% MF coal		
Volatile Matter	42.0	39.6
Fixed Carbon	42.0	38.5
Ash	15.4	21.0
Sulfur	0.6	0.9
Ultimate Analysis, wt% MF coal		
Carbon	61.2	52.1
Hydrogen	4.5	4.0
Nitrogen	1.0	0.5
Sulfur	0.6	0.9
Oxygen	17.3	21.5
Ash	15.4	21.0
Petrographic Analysis, vol%		
Vitrinite*	81.2	77.6
Exinite	7.9	1.7
Inertinite	9.9	18.7
Mineral Matter	0.9	2.0
Ash Analysis, wt%		
Silica, SiO_2	47.1	42.1
Alumina, Al_2O3	17.8	15.6
Titania, TiO_2	**	0.9
Ferric Oxide, Fe_2O_3	2.5	6.7
Lime, CaO	11.7	14.3
Magnesia, MgO	2.9	5.0
Potassium Oxide, K_2O	0.5	1.3
Sodium Oxide, Na_2O	5.0	1.8
Phosphorous Pentoxide, P_2O_5	1.0	12.3
Sulfur Trioxide, SO_3	1.0	
Undetermined	10.5	

* Equivalent to huminite in low-rank coal

** Undetermined

Source: Information package received from Saskoil at the Project Kick-off Meeting, Sept. 10-11, 1981; letter from F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981 to J.A. Mikhlin, SNC Inc.

Petrographic Analysis from University of Regina Report to Saskoil, May 1982.

2.3 OVERALL AND LIQUID PRODUCT EFFICIENCY (cont'd)

where electricity is required from an outside source, a thermal equivalent factor of 10,000 Btu/kWh (10,550 kJ/kWh) as suggested by ESCOE*, rather than the true thermodynamic equivalent factor of 3413 Btu/kWh, has been used to allow realistic conversion loss for electric power generation.

Since the objective of the study is to investigate coal liquefaction processes for the production of transportation fuels, the "liquid" product efficiency is also calculated. Here, the liquid product refers to distillate product in the temperature range of C_5 - 525°C.

2.4 OVERALL PLANT CONFIGURATION, PLANT SIZE AND BATTERY LIMITS

All processes are considered as grass roots facilities. The plant comprises self-sufficient, integrated operation requiring coal, water and electricity (if required) as major raw material and process inputs.

The "as-received" coal supplied to the plant is assumed to be run-of-mine (ROM). No pretreatment/beneficiation of coal to reduce moisture and ash (minerals) contents is considered. Development of coal mine, water supply, gas delivery pipelines and booster stations, product transportation facilities, etc., are also not considered.

Plant Size and Capacity Factor : The conceptual, commercial plant will be based on a feedrate of 25,000 tonnes of as-received coal charged to the plant. The capacity factor of 0.85, i.e., 310 operating days per year is assumed.

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* B.C. McBeath, The Engineering Society Commission on Energy Inc., "Guidelines for Coal Conversion Processes", DOE Report No.FE-2468-44, April 1979.

2.4 OVERALL PLANT CONFIGURATION, PLANT SIZE AND BATTERY LIMITS (cont'd)

Plant Location and Coal Delivery : The plant is assumed to be located within 16 to 24 km (10 to 15 miles) radius of the coal mines. No major site development other than leveling or grubbing has been assumed. Coal is assumed to be delivered to the plant by unit railcars. If, however, the plant site is within 3 km (2 miles), transportation by trucks or conveyors may be economical.

Ash/Slag Disposal : Ash/slag is transported back to the mines for disposal.

Coal Fines : Coal fines may be burned for steam and power generation. Excess coal fines will be returned to the mines without credits.

Water Availability : Water is assumed to be available to the plant site from Diefenbaker Lake through a canal to be built. Supply of water will depend on Provincial Government priorities. It is further assumed that the cost of water will be the same as that for water presently supplied by the Saskatoon South-East Water Supply canal from Diefenbaker Lake to the potash mines and the communities along its route.

Steam and Power Availability : Steam is generated on site by recovering process heat and burning coal. Power is also generated onsite. Additional electricity, if required, will be purchased from the utility network and a conversion factor of 10,000 Btu/kWh (10,500 kJ/kWh) will be applied for thermal efficiency calculations.

2.5 ENVIRONMENTAL ASPECTS

The conceptual plant design will meet all existing local, provincial and federal environmental regulations for particulates, liquid and gaseous effluents.

2.6 ECONOMIC CRITERIA

Preliminary or budget-type capital cost estimate, operating, maintenance and product cost evaluation is carried out for each of the "preselected" coal liquefaction processes only (Stage II of the Study). The basic assumptions and criteria are summarized in Table 2-2.

Each process is evaluated as a self sufficient, integrated, grass-roots facility processing 25,000 tonnes per day of as-received Willow Bunch lignite supplied to the plant. The plant is assumed to be located within 16 km (10 miles) of the coal mines. The feed coal to the plant is assumed to be sized to minus 5-cm and having typical (average) ultimate and proximate analyses as shown in Appendix A.

All costs are calculated in "constant" mid-1982 Canadian dollars and no forward escalation beyond mid-1982 included. Cost of capital is assumed to be 10% per annum. For 100% debt financing this value is equivalent to debt* cost in "real" terms. For 100% equity financing, it may be considered as return on equity before taxes.

Cost estimates and data received from process developers or licensers for a different plant location, capacity and conditions

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* for calculations on "constant" dollar basis, debt cost is expressed in "real" terms, i.e., prevailing interest rates minus inflation.

Table 2-2 Financial Assumptions and Economic Criteria

Coal feed: 25,000 t/d to the plant.

Constant Canadian dollars, mid-1982.

Plant location: within 16 km (10 miles) of the coal mine

4-year construction period.

Capital investment schedule (S-type curve)

Year	%
- 4	9
- 3	25
- 2	36
- 1	30

Investment tax credit for the cost of borrowing during construction claimed during year of occurence.

Plant life: 30 years

Straight-line depreciation method

Zero salvage value

Process contingency: 5%

Project and estimating contingency: 10%

Plant start-up efficiencies

Year	% Design Capacity
1	50
2	90
3	100

Plant service factor: 0.85 (310 operating days per year)

Cost of capital financing = 10% per annum.

Land cost: $990/hectare ($400/acre)

Working capital: 3% of installed-plant cost

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will be converted and adjusted to designated Saskatchewan location.

A 4-year construction period, 30-year operating life, straight-line depreciation and zero salvage value are assumed. The capital investment schedule during construction is based on a typical S-type curve (see Appendix D-1) and the plant is expected to achieve full design capacity in 3 years (see Appendix D-2).

Operating, maintenance and production costs for each of the "preselected" coal liquefaction processes will be reported on an annual basis assuming a plant service factor of 85%, i.e., 310 operating days per year at 100% design capacity (25,000 t/d of coal).

Product costs for preselected coal liquefaction processes will be calculated on "Energy Value" and "Product Value" basis. The first method is based only on the heating value whereas the second also considers the quality of the individual product. The latter approach, Product Value method, was developed by The Engineering Society Commission on Energy Inc. (ESCOE)* and the objective is to allow relative comparison and ranking of the energy related processes.

Byproduct credits for sulfur at $65/t, NH_3 at $300/t and mixed phenols at $250/t have been taken into consideration.

Large quantity of CO_2 produced in both direct and indirect coal liquefaction processes have a potential for utilization in enhanced oil recovery operation. No credits are taken for CO_2 byproduct, since the market for such a use is quite uncertain.

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* K.A. Roger and R.F. Hill, "Coal Conversion Comparison", DOE Report No. FE-2468-51, July 1979.

DIRECT COAL LIQUEFACTION:

3.0 OVERVIEW OF ADVANCED COAL LIQUEFACTION PROCESSES AND THEIR SUITABILITY FOR DESIGNATED SASKATCHEWAN LIGNITES

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3.0 DIRECT COAL LIQUEFACTION: OVERVIEW OF ADVANCED COAL LIQUEFACTION PROCESSES AND THEIR SUITABILITY FOR DESIGNATED SASKATCHEWAN LIGNITES

In this summary volume (Volume I), a brief overview of the five advanced "direct" coal liquefaction processes is presented along with the specific features of each process, its status, and its suitability for the designated Saskatchewan coals (Estevan and Willow Bunch lignites). A comprehensive technical assessment including process history and development, relevant experimental results pertaining to processing of lignites, overall configuration of a conceptual commercial plant, problem areas and commercialization constraints, etc., are described in Volume II.

The advanced direct coal liquefaction processes presented in this chapter are considered for commercialization by 1990. The processes selected by the Study Committee are:

- H-Coal, Hydrocarbon Research Inc., U.S.A.
- Exxon Donor Solvent (EDS), Exxon Research and Engineering Company, U.S.A.
- Modified I.G. Farben (New German Technology for High-Rank Coals), Ruhrkohle AG and Veba Oel AG, F.R.G.
- Modified I.G. Farben (New German Technology for Low-Rank Coals), Rheinische Braunkohlenwerke AG, F.R.G.
- Liquid Solvent Extraction (LSE), National Coal Board, U.K.

Comparative evaluation of these direct coal liquefaction processes have been carried out in this chapter for prescreening and ranking for Saskatchewan lignites (Stage I of the Study). The selected direct coal liquefaction process has been considered for further techno-economic assessment and compared with the selected indirect coal liquefaction processes in Chapter 6.

3.1 H-COAL PROCESS

The H-Coal process is a direct, catalytic hydroliquefaction process developed by Hydrocarbon Research Inc. (HRI) to convert coal into boiler fuels (fuel-oil mode) or synthetic crude oil (syncrude mode). The H-Coal process is a novel application of the commercially proven "ebullated-bed" reactor originally employed by HRI to convert heavy oil residues into lighter fractions (H-Oil process).

3.1.1 Process Description

A simplified flow diagram of the H-Coal process is given in Figure 3-1. Coal is pulverized, dried and then mixed with recycle oil to form a coal/oil slurry containing approximately 50% solids by weight. The mixture is preheated and fed into an ebullated-bed catalytic reactor where the coal is liquefied at high temperature (425-455°C) and pressure (15-21 MPa) in the presence of hydrogen. The reactor effluents leaving from the top of the reactor are cooled and then separated into gases, liquids and a residual slurry by a conventional flash separation, hydroclones, and vacuum distillation. The catalyst being heavier than the products is retained in the reactor. The spent catalyst is periodically withdrawn and replaced by fresh catalyst through the on-line "catalyst addition and withdrawal system". The primary products from the process are LPG, naphtha (C_5-200°C), mid-distillate (200-345°C) and heavy oil (345-525°C). Byproducts include sulfur, ammonia and crude phenols.

Make-up hydrogen is produced by gasification of residual slurry (vacuum bottoms) in Texaco's partial oxidation unit, followed by shift conversion and gas treatment.

3.1 H-COAL PROCESS (cont'd)

3.1.1 **Process Description** (cont'd)

For a conceptual plant processing 25,000 t/d of designated Saskatchewan lignites, the estimated make-up water requirement is 360 L/s. The electricity consumption is about 180 MW but all the power is generated on site.

3.1.2 **Product Yields and Product Efficiencies**

From coals tested in the H-Coal process, bituminous Illinois No.6 (27,600 hours of operation) and subbituminous Wyodak (12,672 hours of operation) have been investigated most extensively in the bench-scale, PDU and the pilot plant. North Dakota (550 hours of operation) and Texas lignites (705 hours of operation) have been tested for a short period over 15 years ago and the results do not represent the present stage of development of the H-Coal process. Typical analyses of these coals are given in Appendix B.

Comparative product yield and distribution for a conceptual, self-sufficient, commercial H-Coal plant under syncrude mode of operation with bituminous Illinois No. 6, subbituminous Wyodak, Estevan and Willow Bunch lignites are presented in Table 3-1. For Saskatchewan lignites, the product yields were estimated on the basis of HRI's PDU Run No. 6 and conceptual plant data for Wyodak coal (EPRI Report No. AF-1143-SR, July 1979).

The results show that the net liquid-product (C_5-525°C) yield decreases from 2.6 bbl/t for bituminous coal to 1.5 bbl/t for Estevan lignite and to 1.0 bbl/t for Willow Bunch lignite on an as-received basis. In other words, the yield is about 30% lower for Wyodak coal, 40% lower for Estevan, and 60% lower for Willow Bunch lignite as compared to bituminous Illinois No. 6 coal.

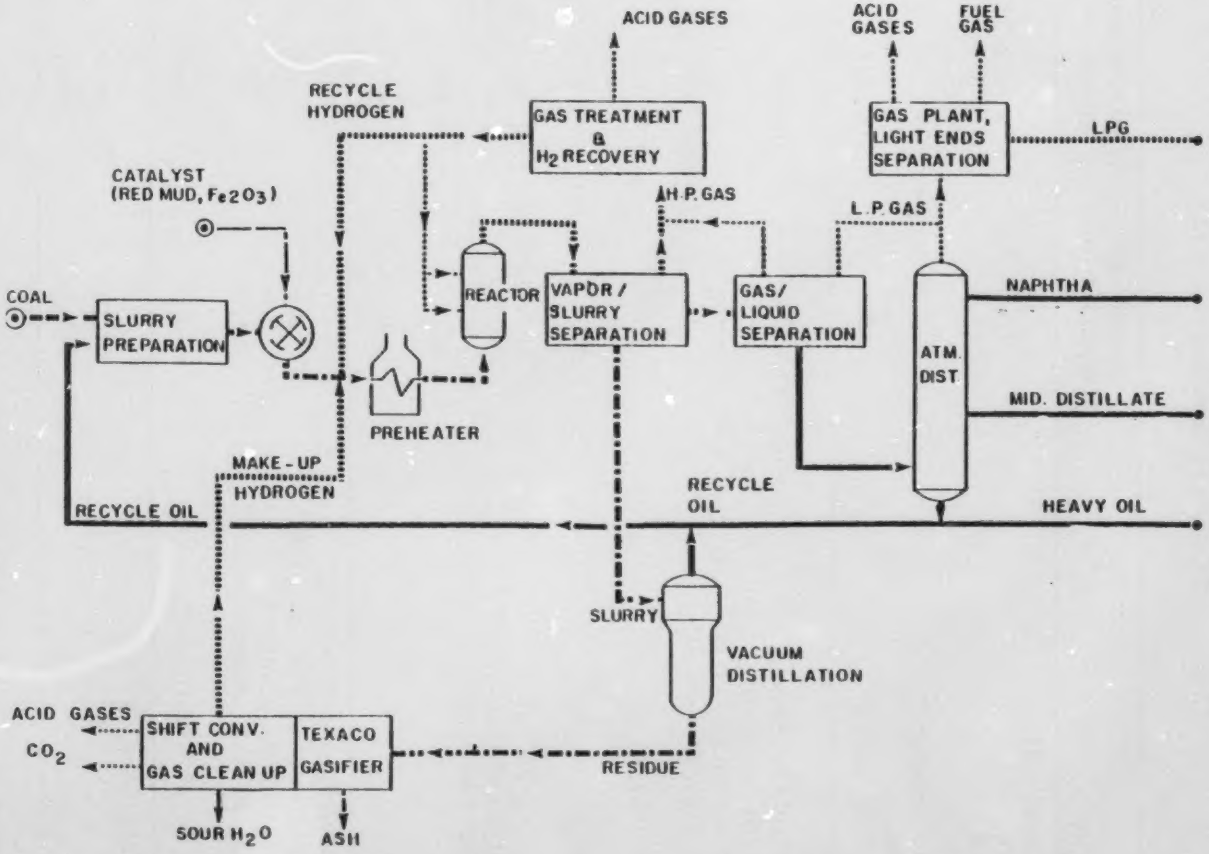

Figure 3-1 Overall Conceptual Configuration of the H-Coal Process

Table 3-1 Product Yield and Efficiencies of the Conceptual H-Coal Plant

[Basis: 25,000 t/d as-received coal]

	Bituminous Illinois No.6	Subbituminous Wyodak
Coal (as received)		
Moisture, wt%	10.0	30.0
Ash, wt%	10.6	8.6
Coal Fed to Reactor		
Moisture, wt%	2.0	2.0
Ash, wt% MF coal	10.6	8.6
Input		
MF Coal to the Plant, t/d	22,500	17,500
MF Coal to Gasifiers, t/d	-	-
MF Coal to Utilities and Offsites, t/d	-	-
Electricity, MW	-	-
Ouput, t/d (bbl/d)		
SNG	-	-
Propane	330	-
Butane	534	-
Naphtha (C_5-200°C)	2,807 (22983)	2,708 (21293)
Mid-distillate (200-345°C)	4,170 (27862)	2,327 (16500)
Heavy Oil (345-525°C)	2,576 (13035)	1,440 (9132)
Sulfur	767	153
Phenol	62	20
Ammonia	256	141
Liquid Product yield (C_5-525°C)		
bbl/t as-received coal to plant	2.6	1.9
bbl/t MF coal to plant	2.9	2.7
bbl/t MAF coal to plant	3.2	3.0
Naphtha in Total Liquid, wt%	30.0	41.0
Product Efficiency, %		
Overall	71.0	62.0
Liquid	62.0	61.0

Sources: EPRI Reports No. AF-1143-SR, July 1979; and AF-1297, Dec. 1979

Table 3-1 Product Yield and Efficiencies of the Conceptual H-Coal Plant (cont'd)

[Basis: 25,000 t/d as-received coal]

	Estevan Lignite	Willow Bunch Lignite
Coal (as received)		
Moisture, wt%	35.0	35.0
Ash, wt%	15.4	21.0
Coal Fed to Reactor		
Moisture, wt%	2.0	2.0
Ash, wt% MF coal	15.4	21.0
Input		
MF Coal to Reactor, t/d	16,250	14,428
MF Coal to Gasifiers, t/d	-	-
MF Coal to Utilities, t/d	-	1,822
Electricity, MW	-	-
Ouput, t/d (bbl/d)		
SNG	-	-
Propane	50	-
Butane	82	-
Naphtha (C_5-200°C)	2,327 (18300)	1,420 (11164)
Mid-distillate (200-345°C)	2,000 (14180)	1,660 (11768)
Heavy Oil (345-525°C)	865 (5488)	286 (1813)
Sulfur	87	142
Phenol	17	20
Ammonia	145	88
Liquid Product Yield (C_5-525°C)		
bbl/t as-received coal to plant	1.52	0.99
bbl/t MF coal to plant	2.34	1.52
bbl/t MAF coal to plant	2.76	1.93
Naphtha in Total Liquid, wt%	45.0	42.0
Product Efficiency, %		
Overall	62.0	46.8
Liquid	59.0	45.7

3.1.2 **Product Yields and Product Efficiencies** (cont'd)

This is due to high moisture and ash contents of the low-rank coal and because the higher-rank coals generally give higher product yields. This may be evident by observing the net-liquid yields expressed in terms of moisture-free (MF) and moisture-and-ash free (MAF) basis. On an MAF basis, the net-liquid yield is only 10% lower for Wyodak coal and about 16% lower for Estevan lignite compared to Illinois No.6 coal. For Willow Bunch lignite, the product yields are lower because of higher ash content (21 wt% on MF basis) and correspondingly greater loss of recoverable liquid in the residual slurry.

The overall product efficiency is in the range of 47% to 71% depending on the coal feedstock. The liquid product efficiency for Estevan and Willow Bunch lignites is 59% and 46% respectively.

3.1.3 **Process Status**

The H-Coal process has been under development at HRI since 1964. More than fourteen types of coal have been evaluated in over 68,000 hours of operation. The development and demonstration of the process have been carried out in micro-autoclaves, bench-scale units (11 kilograms per day of coal) and a process development unit (PDU) having a capacity of processing up to 3.2 tonnes per day of coal. Recently, the feasibility of the process was demonstrated in a pilot plant located at Catlettsburg, Kentucky. The pilot plant has a capacity of processing up to 225 tonnes per day of coal in the syncrude mode and up to 545 tonnes per day of coal in the fuel-oil mode.

3.1.3 Process Status (cont'd)

A detailed design study for a commercial H-Coal plant to be located at Breckinridge County in Kentucky was initiated under the sponsorship of the U.S. Department of Energy as a cooperative venture with Airco Energy Co. and Ashland Oil Inc. In October 1981, Bechtel joined in as a partner in the Breckinridge project and will be responsible for process design and engineering but recently Airco Energy Co. announced its withdrawal from the project due to financial constraints. A proposal for U.S. government support has been submitted to the U.S. Synthetic Fuels Corporation. The current schedule calls for detailed engineering design to begin in 1982 leading to construction in 1983 with initial production in 1988.

3.1.4 Suitability of Saskatchewan Lignites

Although the process is flexible in terms of coal feed, only two experimental runs have been made with lignites in a bench-scale unit. In 1966, North Dakota lignite from the Dickinson area was tested (550 hours of operation) when the program was under the sponsorship of the Office of Coal Research, U.S. Department of Interior (now U.S. Department of Energy). A second experimental run (705 hours of operation) was made in 1967 with Texas lignite for a private client. Recently, a high ash Australian brown coal from Muswellbrook has been tested by HRI* in a micro-autoclave unit to determine its coal liquefaction behaviour in the H-Coal process. The results of the tests were projected for a conceptual, commercial plant evaluation using a computer program by HRI for The Energy Authority of New South Wales, Australia (HRI Report, Aug. 1981). The ash content of this coal was assumed to

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* J.L. Gendler, HRI, Letter to J.A. Mikhlin, SNC Inc., HRI Ref. No. 10-JLG-165-81, October 2, 1981 (Appendix C-1).

3.1.4 **Suitability of Saskatchewan Lignites** (cont'd)

be reduced to 10 wt% by beneficiation prior to feeding into the coal liquefaction plant.

In evaluating the suitability of Estevan and Willow Bunch lignites, certain inherent characterics such as high ash and moisture contents which affect their behaviour in coal liquefaction process must be considered.

As-received Estevan and Willow Bunch lignites contains 35.0 wt% moisture which must be reduced to between 2% and 4 wt% before feeding the coal in the reactor. Excess moisture requires higher total operating pressure inside the reactor to maintain a given partial pressure of hydrogen. In addition, presence of moisture results in emulsion formation and may create foaming problems in the coal/oil slurry preparation unit.

Drying of low-rank coals with air has been observed to reduce their activity with respect to liquefaction, possibly because of surface oxidation and collapsing of pore structure. For Estevan and Willow Bunch lignites, this aspect should be experimentally investigated.

The Estevan lignite has 15.4 wt% ash on an MF basis. In Willow Bunch lignite, the ash content is even higher (21 wt% ash). Consequently, the vacuum bottoms (residual slurry) would have a high viscosity due to high solid-content* and low residual-oil (> 525°C) yield. This affects the operability of the vacuum distillation tower and subsequent processing units due to

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* Solids in residual slurry includes unreacted coal, ash and inertinite matter of the coal and should typically not exceed 50 wt%.

3.1.4 Suitability of Saskatchewan Lignites (cont'd)

increased difficulty in pumping the vacuum bottoms. In the conceptual plant configuration for Estevan and Willow Bunch lignites, dilution of residual slurry with recoverable heavy oil (345-525°C) was considered as a means of controlling the vacuum-bottoms viscosity. The net liquid (C_5-525°C) yield dropped by 0.2 bbl/t of MF coal for Estevan lignite and 0.5 bbl/t for Willow Bunch lignite. In processing of Willow Bunch lignite, the loss of liquid product is high and problems associated with handling of residual slurry may be acute.

Catalyst consumption (without regeneration) in the H-Coal reactor is typically in the range of 0.3 to 1.2 kg/t MAF coal depending on the type of catalyst used, type of coal processed, severity of operating conditions employed and quality of final product required. In the conceptual plant configuration of the H-Coal process, standard cobalt molybdate catalyst (American Cynamid HDS-1442A) was considered and a catalyst replacement rate of 0.5 kg/t MF coal (or about 0.6 kg/t MAF coal) recommended by HRI was taken. This value has been established during extensive PDU runs and has also been verified in the recent pilot-plant operation. At this rate, the annual catalyst consumption will be approximately 2.6 Gg (2600t). The supply of catalyst in such an amount is a major concern, especially the availability of cobalt.

In conclusion, the main concerns for processing of Saskatchewan lignites in the H-Coal process would be due to its high ash and moisture contents resulting in lower liquid product yields and thermal efficiencies. Catalyst consumption may also be high due to rapid poisoning of catalyst by mineral matter present in ash.

3.1.4 **Suitability of Saskatchewan Lignites** (cont'd)

Also, processibility and handling of solids-containing slurry (vacuum bottoms) could be a problem. Furthermore, in a separate evaluation for Muswellbrook coal (22.6 wt% ash MF coal), HRI concluded that their H-Coal process probably would not be feasible for commercial coal liquefaction unless the ash content is economically reduced to 10 wt% by coal pretreatment/beneficiation (for details, see Appendix C-1).

3.2 RUHRKOHLE/VEBA OEL PROCESS

Ruhrkohle AG and Veba Oel AG are developing a direct, catalytic coal liquefaction process for the production of transportation fuels and/or petrochemical feedstocks. The RAG/Veba process, often referred to as "New German Technology" is a modified and improved version of the classical Bergius/I.G. Farben approach.

In the RAG/Veba process, several changes and improvements have been incorporated over the classical Bergius/I.G. Farben process, some of them are mentioned below:

- Vacuum distillation and flashing methods are used for solid/liquid separation instead of centrifuging and coking.
- Asphaltene-free distillate oil is used as recycle solvent instead of residual oil containing asphaltenes and solid residues.
- Partial oxidation of vacuum bottoms (residues) to produce make-up hydrogen instead of gasification of coke purchased from an outside source.

These improvements have resulted in reduction of reactor operating pressure from 70 MPa to 30 MPa; improved heat recovery and thermal efficiency and lower hydrogen consumption.

3.2.1 **Process Description**

A simplified process flow diagram is presented in Figure 3-2. Dried and pulverized coal (1 wt% moisture, minus 1-mm size) is mixed with recycle solvent (medium/heavy oil: 40/60) and the

3.2.1 **Process Description** (cont'd)

slurried coal is further ground to about 0.2-mm size. During secondary grinding, the required amount of catalyst (6.5 wt% of MAF coal) is also added to form a coal/oil/catalyst slurry containing about 40 wt% solids. The catalyst used is red mud, a waste clay obtained from bauxite processing, containing 30 to 40 wt% Fe_2O_3 as catalytically active component. The resulting slurry is preheated to 425°C and mixed with make-up hydrogen before feeding into the reactor operating at a pressure of about 30 MPa. In the reactor, an exothermic reaction is carried out but the temperature is maintained at 475°C by injecting cold hydrogen at different levels in the reactor. The effluents are cooled and separated into gases, liquid and residual oil products by conventional flash separation and vacuum distillation techniques. The primary products are LPG, naphtha (C_5 - 200°C), mid-distillate (200 - 325°C) and heavy oil ($>$325°C). Byproducts include sulfur, ammonia and crude phenols.

Make-up hydrogen is produced by gasification residual slurry (vacuum bottoms) in Ruhrchemie/Ruhrkohle/Texaco gasifier configuration.

For a conceptual plant processing 25,000 t/d of designated Saskatchewan lignites, the estimated make-up water requirement is 400 L/s. The electricity requirement from an outside source is 160 MW. Catalyst consumption is 6.5 wt% of MAF coal fed to the reactor amounting to 213,200 tonnes/year of red mud.

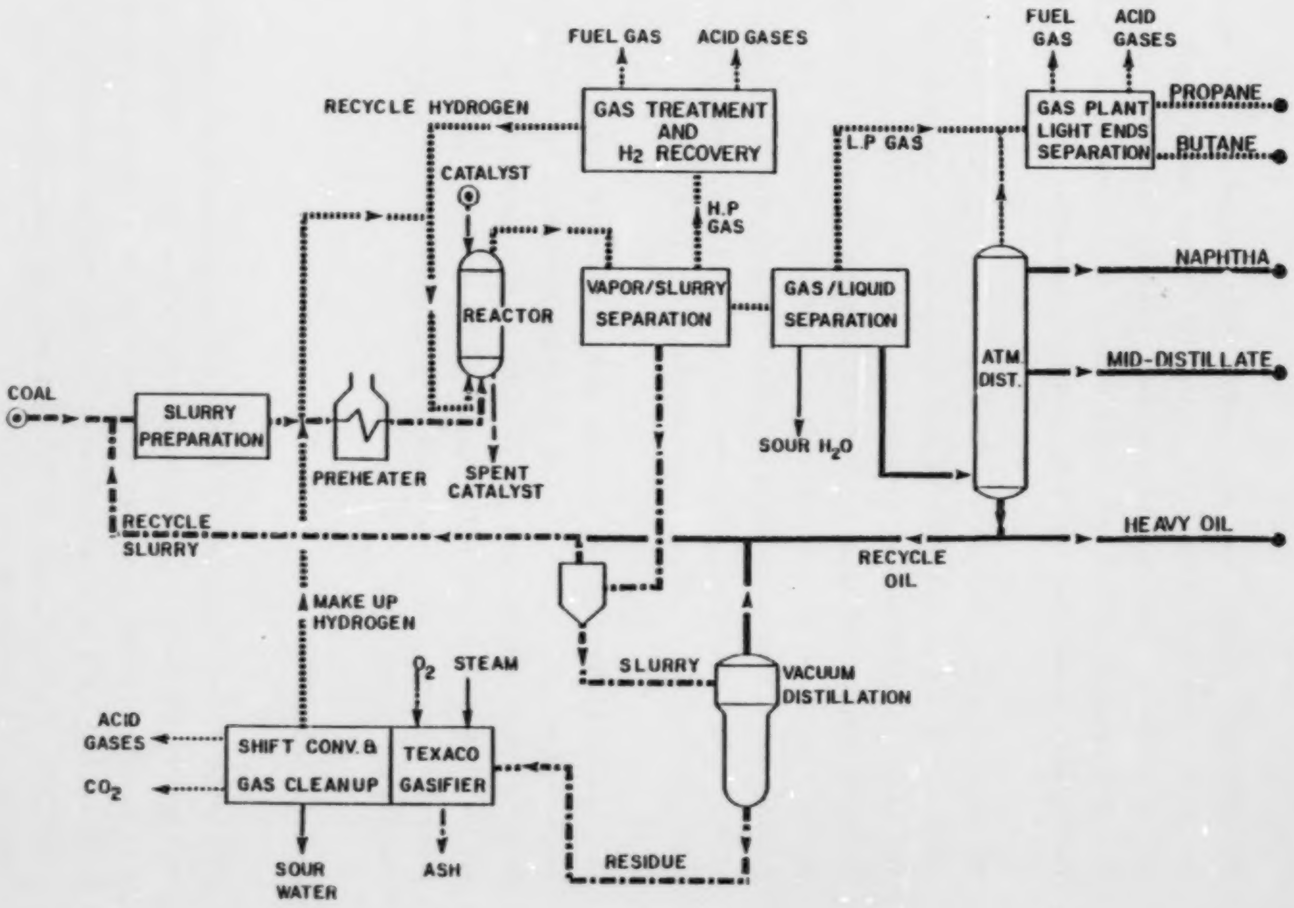

Figure 3-2 Overall Conceptual Configuration of the RAG/Veba Process (New German Technology)

3.2.2 Product Yields and Product Efficiencies

The RAG/Veba process has been extensively tested in the bench-scale unit and the pilot plant using bituminous German coal from the Ruhr area. Analysis of the Ruhr coal is given in Appendix B. The product yield and distribution of a conceptual, self-sufficient, commercial RAG/Veba plant is presented in Table 3-2.

The results show that the net liquid product (C_5-525°C) yield is about 2.3 bbl/t of as-received bituminous German coal compared to 1.0 bbl/t of as-received Estevan lignite. In other words, the yield is almost 40% less for Estevan lignite and this is because of its high moisture and ash content, and also because the bituminous coals generally give higher product yields. This is evident by observing the net liquid yields expressed in moisture-free (MF) and moisture-and-ash-free (MAF) basis.

The "overall" product efficiency for bituminous German coal is 60.5% versus 55.7% for Estevan lignite. The "liquid" product efficiencies are 49.1% and 44.0% respectively. For Willow Bunch lignite with higher ash content, even further reduction in product yields and efficiencies may be expected in comparison to bituminous German coal.

3.2.3 Process Status

Ruhrkohle AG and Veba Oel AG have been developing their modified version of Bergius/I.G. Farben process since 1974. During this period, the process has been demonstrated on a bench-scale unit (250 to 500 kg/d of coal) located at the Bergbau-Forschung GmbH facilities at Essen, and in 120 kg/d unit at the laboratories of Veba Oel AG at Scholven.

Table 3-2 Product Yield and Efficiencies in the Conceptual RAG/Veba Plant

[Basis: 25,000 t/d as-received Coal]

	Bituminous German Coal	Estevan Lignite
Coal (as received)		
Moisture, wt%	4.3	35.0
Ash, wt% of MF coal	4.5	15.4
Coal Fed to Reactor		
Moisture, wt%	1.0	2.0
Ash, wt% of MF coal	4.5	15.4
Input		
MF Coal to Reactor, t/d	21165	12505
MF Coal to Gasifiers, t/d	-	-
MF Coal to Utilities, t/d	2759	3745
Electricity, MW	230	160
Output, t/d (bbl/d)		
SNG	-	-
Propane/Butane (LPG)	1818	952
Naphtha (C_5-200°C)	2929 (20960)	1058 (7570)
Mid-Distillate (200-325°C)	6972 (44290)	3553 (22570)
Heavy Oil (325-525°C)	201 (1200)	-
Sulfur	81	57
Phenol	-	-
Ammonia	263	116
Liquid Product Yield, (C_5-525°C)		
bbl/t as-received coal to plant	2.3	1.0
bbl/t MF coal to reactor plant	2.4	1.5
bbl/t MAF coal to reactor plant	2.5	1.7
Naphtha in Total Liquid, wt%	29.0	23.0
Product Efficiency, %		
Overall	60.5	55.7
Liquid	49.1	44.0

Note: Product yields and efficiencies for both bituminous German coal and Estevan lignite have been adjusted to zero electricity import using a factor of 10,000 Btu/kWh (10,550 kJ/kWh) and a fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

Source: Conceptual plant data (19100 short ton/day of coal) and information package supplied by Veba Oel AG, Ref. No. 5S3/hs-ku, 27 July 1981.

3.2.3 **Process Status** (cont'd)

Different types and ranks of coal have been tested, but extensive bench-scale tests have been conducted only with bituminous German coal from the Ruhr area (referred to as "standard" coal by process developer).

On the basis of the bench-scale results and the past operating experience with the old I.G. Farben plants in Germany, Ruhrkohle AG and Veba Oel designed, constructed, and are operating a 200-t/d pilot plant at Bottrop. This plant was directly scaled up from a 0.5-t/d bench-scale unit and is in operation since March 1982. A three-year operation is planned and one of the major objectives of the pilot plant program is to demonstrate the long-duration and trouble-free operation of the process.

The final objective of the process development program is to advance the RAG/Veba coal liquefaction technology towards commercialization in the 1990. Two independent proposals have been submitted by Ruhrkohle AG and Veba Oel AG to the Federal Ministry of Research and Technology in Bonn, namely:

- Ruhrkohle AG: Commercial plant (6 million tonnes per year of coal) for the production of 2 million tonnes of primary liquid products from bituminous coal at a site in the Ruhr area. The byproducts are 0.6 million tonnes of LPG and about 1.8 billion cubic metres of SNG.

- Veba Oel AG: Commercial plant processing about 6 million tonnes per year of coal, or oil residues, or coal and oil residues to produce about 2 million tonnes per year of primary

3.2.3 **Process Status** (cont'd)

liquid products. The Veba Oel AG concept to use the same plant configuration for processing either coal or heavy oil or residues (but not in combination) allows more operating flexibility with changing market demand and economic requirements.

So far, no decision has been taken to proceed with these commercial plants. However, on the basis of past operating experience with I.G. Farben plants and present state of development, it may be expected that the RAG/Veba process would reach commercial readiness by 1990.

3.2.4 **Suitability of Saskatchewan Lignites**

To date, extensive testing in a bench-scale unit and the pilot plant have been carried out using bituminous German coal. No tests have been made on lignites.

In evaluating the suitability of Saskatchewan lignites for the RAG/Veba process, similar areas of concern outlined earlier for H-Coal process such as lower product yields, lower thermal efficiency, and difficulty in handling of vacuum bottoms will apply. However, in the case of RAG/Veba process, the problems are more severe because of the type of catalyst used. The red mud containing 30 to 40% Fe_2O_3 as catalyst is readily available and inexpensive but the amount of red mud used is 6.5 wt% MAF coal and all of it eventually ends up in the residual slurry (vacuum bottoms). Consequently, the vacuum bottoms (residues) have a higher viscosity due to high solid content (ash, unreacted coal and catalyst). This would affect the operation of the

3.2.4 **Suitability of Saskatchewan Lignites** (cont'd)

slurry handling equipments such as slurry pumps, vacuum distillation and subsequent processing units due to increased pumping difficulty.

The net liquid product (C_5 - 525°C) yield on an MAF basis for Estevan lignite (1.5 bbl/t MF coal) is much lower than that for bituminous German coal (2.4 bbl/t MF coal) as shown in Table 3-2. This is because a large amount of recoverable liquid product is added to the residual slurry to make it pumpable. For Willow Bunch lignite (21 wt% ash MF coal), the net liquid product yield would be even lower.

The above areas of concerns in the RAG/Veba process for Saskatchewan lignites have also been pointed out by Veba Oel AG (see Appendix C-2). These problems will prevail unless the ash content of Saskatchewan lignites can be economically reduced to less than 10% by coal pretreatment/beneficiation. Ruhrkohle/Veba Oel considers their process to be uneconomical for Saskatchewan lignites.

3.3 RHEINBRAUN PROCESS

Rheinische Braunkohlenwerke AG (Rheinbraun) in collaboration with its subsidiary Union Rheinische Braunkohlen Kraftstoff AG is developing a direct, catalytic process for liquefaction of German brown coal.

In the Rheinbraun process, a number of modifications have also been made over the classical Bergius/I.G. Farben approach such as use of asphaltene-free distillate oil as recycle solvent, vacuum distillation and flashing methods for solid/liquid separation, and gasification of vacuum bottoms (residues) to produce make-up hydrogen for the process. These changes have resulted in increased coal throughput from 400 kg/h/m^3 to 600 kg/h/m^3, lowering of reactor pressure from 71 to 30 MPa, reduction of reactor temperature from 478° to 460°C, and lower hydrogen consumption (6.0 wt% instead of 6.5 wt% MAF coal).

While these improvements are similar to the RAG/Veba process described earlier, the Rheinbraun process incorporates some processing steps specifically for brown coals, such as:

- A proprietary method for drying of raw brown coal containing up to 60 wt% moisture (as-received coal).

- Thermal pretreatment of coal/oil/catalyst slurry at 350°C to remove residual moisture and to decompose the humic acids which create calcium-carbonate induced deposits inside the preheater tubes and hydrogenation reactor.

3.3 RHEINBRAUN PROCESS (cont'd)

3.3.1 **Process Description**

Figure 3-3 is a simplified block diagram of the Rheinbraun process. The raw coal from storage bins is fed into the crusher/dryer where the coal is pulverized to minus 0.1 mm and dried to between 10 and 12 wt% moisture. A major portion of the coal (67% of the total) is fed into a slurry preparation unit in the coal liquefaction section. The balance is used for hydrogen, fuel gas, steam and power generation.

The partially dried and pulverized coal is mixed with recycle oil (MF-coal/oil ratio of 0.75) and red mud (5 wt% MAF coal) to form a slurry containing 45 wt% solids. In the slurry preparation vessel, the temperature is maintained at 350°C to remove the residual moisture and to thermally decompose the humic acids present in low-rank coals.

The effluents leaving the reactor are separated into gases, liquid and slurry streams by conventional flash separation and vacuum distillation methods. The primary products are LPG, naphtha (C_5-200°C), mid-distillate (200 -325°C) and heavy oil (> 325°C). Byproducts include sulfur, ammonia and phenols.

Make-up hydrogen is produced by gasification of residual slurry (vacuum bottoms), followed by shift conversion and gas treatment.

For a conceptual plant processing 25,000 t/d of Saskatchewan lignites, the estimated make-up water requirement is about 525 L/s. Catalyst consumption is 5 wt% MAF coal amounting to 168,330 tonnes per year of red mud. Electricity is generated on site.

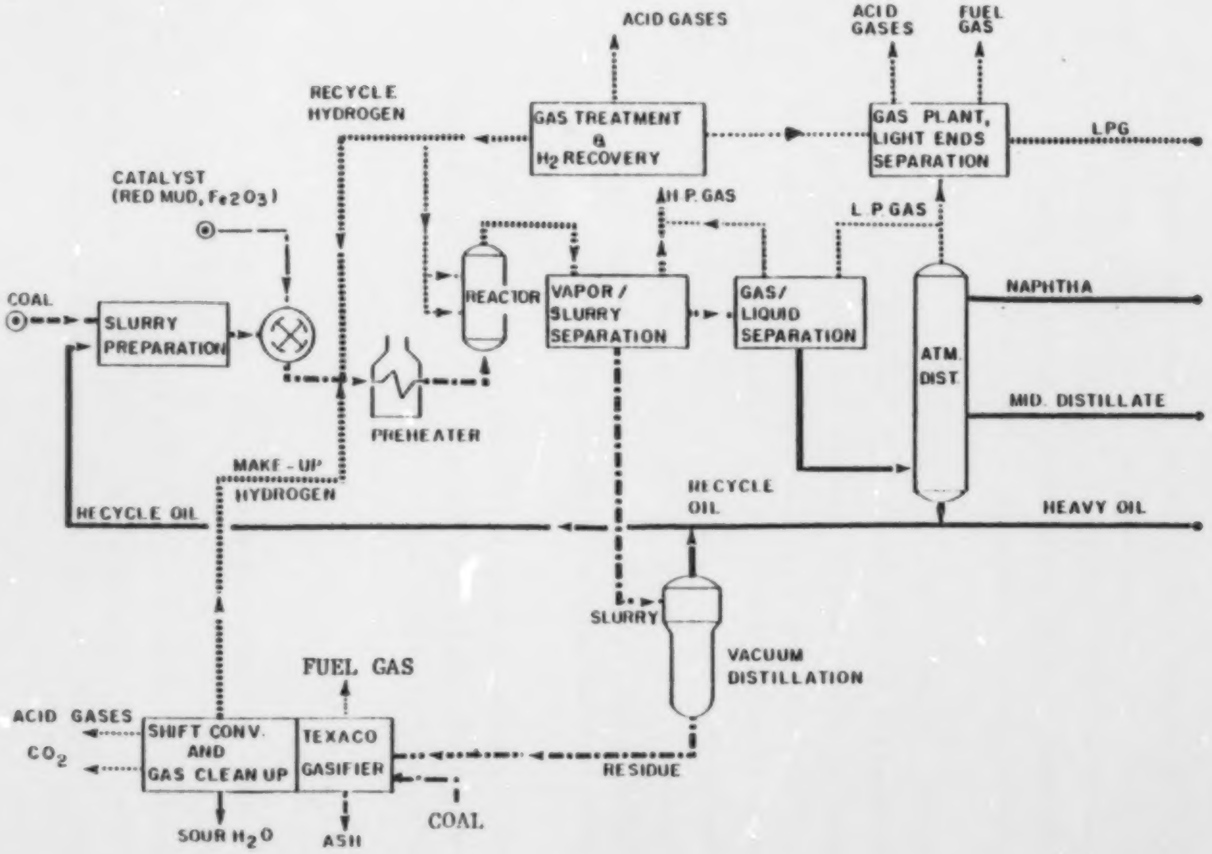

Figure 3-3 Overall Conceptual Configuration of the Rheinbraun Process

3.3.2 Product Yields and Efficiencies

The Rheinbraun process has been developed specifically for high moisture but low-ash Rhenish brown coal. Analysis of Rhenish coal is given in Appendix B. Product yield and distribution of a conceptual, self-sufficient, commercial plant of Rheinbraun process plant is presented in Table 3-3.

The net liquid product (C_5-325°C) yield is about 0.7 bbl/t of as-received Rhenish brown coal compared to 1.1 bbl/t of Estevan lignite. The yield on as-received basis for Rhenish brown coal is 36% less because of its high moisture content (60 wt% H_2O). On an MF basis, the yield for Estevan lignite is lower than for Rhenish brown coal because of higher ash content of Estevan lignite. The yields on MAF basis are almost the same. For Willow Bunch lignite, the liquid product yield on MAF basis would be lower because of higher ash content.

The "overall" product efficiency for Rhenish brown coal is 51% compared to 52.2% for Estevan lignite. The "liquid" product efficiency are 44.1% and 43.2% respectively.

3.3.3 Process Status

The new Rheinbraun process has been under development since 1978. The process has been tested with a bench-scale unit having a capacity of processing up to 250 kg/d of coal. Most of the coal liquefaction experiments have been carried out on Rhenish brown coal. Sufficient new experimental data supplemented with past

Table 3-3 Product Yield and Efficiencies of the Conceptual Rheinbraun Plant

[Basis: 25,000 t/d as-received coal]

	Rhenish Brown Coal	Estevan Lignite
Coal (as received)		
Moisture, wt%	60.0	35.0
Ash, wt% of MF coal	3.2	15.4
Input		
MF Coal to Reactor, t/d	6535	10860
MF Coal to Gasifiers* t/d	660	1555
MF Coal to Utilities**, t/d	2805	3835
Electricity, MW	-	-
Output t/d (bbl/d)	-	-
SNG	-	-
LPG	253	592
Naphtha (C_5-200°C)	756 (5950)	1323 (10410)
Mid-Distillate (200-325°C)	1727 (12278)	2397 (17040)
Heavy Oil (325-525°C)#	-	-
Sulfur	27	56
Phenol	-	-
Ammonia	40	76
Liquid Product Yield (C_5-525°C)		
bbl/t as-received coal to plant	0.7	1.1
bbl/t MF coal to plant	1.8	1.7
bbl/t MAF coal to plant	1.9	2.0
Naphtha in Total Liquid, wt%	30.0	35.6
Product Efficiency, %		
Overall	51.0	52.2
Liquid	44.1	43.2

* Coal for fuel gas and hydrogen production

** Coal for steam and power generation

# No heavy oil (325-525°C) is produced since most of it is recycled to the reactor and the balance is added to the residual slurry to make it pumpable.

Source: D. Bocker et al., "Rheinbraun Liquefaction of Brown Coal", paper presented at EPRI/KFA-Julich Conference on Synthetic Fuels at San Francisco, California, October 1980.

3.3.3 **Process Status** (cont'd)

operating experience (1941-44) of the large-scale plants has been accumulated to permit the scale-up of new Rheinbraun technology to the pilot-plant scale (360 t/d of raw brown coal).

Since 1980, Rheinbraun is engaged in the detailed engineering and process design of a 360-t/d pilot plant. The mechanical completion of the pilot plant and the start-up operation are scheduled for 1985 but the target appears to be optimistic. The ultimate objective is to build a single-train commercial plant (annual raw brown coal throughput of 3.0 million tonnes) by 1992 and a full-scale plant (annual raw brown coal throughput of 20 million tonnes) by 1995.

3.3.4 **Suitability of Saskatchewan Lignites**

The new Rheinbraun process has been developed specifically for liquefaction of brown coal from the Rhein area, Germany. Low-rank coals from other countries such as Australia and New Zealand have also been tested.

The as-received Rhenish brown coal contains 60 wt% moisture and has an ash content of 3.2 wt% on MF basis. In comparison, the Estevan and Willow Bunch lignites contain lower moisture (35 wt%) but the ash contents are higher, i.e., 15.4 wt% for Estevan and 21 wt% for Willow Bunch lignites.

The initial moisture content (35 wt%) of the Estevan and Willow Bunch lignites may be reduced to the desired level ($\simeq$ 2 wt%) using a two-stage drying technique suggested by the process developer. Initially, the moisture content is reduced to about

3.3.4 **Suitability of Saskatchewan Lignites** (cont'd)

10 wt% by air drying in the coal preparation section and the residual moisture is subsequently removed during the "thermal pretreatment" step in the slurry preparation unit. The second stage employs a proprietary slurry-drying method, details of which are not available.

Reduction of moisture for Saskatchewan lignites to the desired levels is not expected to be a problem since Rheinbraun has acquired extensive experience and have demonstrated that brown coals containing up to 60 wt% moisture may be dried without significantly affecting the properties of the feed coal and its product yield.

The Estevan and Willow Bunch lignites contain 15.4 wt% and 21 wt% ash respectively compared to 3.2 wt% in Rhenish brown coal. The high-ash content of Saskatchewan lignites together with the addition of red mud catalyst ($\simeq$ 5 wt% of MAF coal feed) to the reactor would contribute to operational difficulties associated with the pumping and handling of the highly viscous and abrasive solids-containing residual slurry.

In the conceptual plant configuration for Estevan lignite, about 4290 bbl/d of recoverable oil was added to the vacuum-bottoms stream (residual slurry) so as not to exceed the limit of 50 wt% solids in the slurry. The net liquid yield decreased by 0.4 bbl/t of MF coal, and the "liquid" product efficiency dropped from 49% to 43%. For Willow Bunch lignite having 21 wt% ash (MF basis), the decrease in net liquid yield would be even more than that for the Estevan lignite.

3.3.4 **Suitability of Saskatchewan Lignites** (cont'd)

The above areas of concern such as low product yields and operational difficulty associated with high-ash content of the coal have also been pointed out by the process developer of the Rheinbraun process (see Appendix C-3) who considers their process as unsuitable for liquefaction of Saskatchewan lignites.

3.4 EXXON DONOR SOLVENT PROCESS

The Exxon Donor Solvent (EDS) coal liquefaction process is a non-catalytic solvent extraction process being developed by Exxon Research and Engineering Co. (ER&E). The main feature of the process is the use of the process-derived hydrogen donor solvent which donates hydrogen to the coal during liquefaction. The donor solvent is a 205 -545°C boiling-range hydrocarbon liquid which is hydrogenated separately in a fixed-bed catalytic reactor prior to recycling to the coal liquefaction reactor.

3.4.1 **Process Description**

Figure 3-4 is a simplified process flow diagram of the EDS process (Design Update Case*). Dried and pulverized coal (4 wt% moisture, 90% minus 2.5 mm) is slurried with recycle solvent and fed to the liquefaction reactor operating at 449°C and 13.8 MPa. The effluents from the extraction reactor are separated into gases, liquid and slurry streams by flash separation methods.

The liquid and slurry streams are fed into the atmospheric distillation unit to recover distillate products. A sidestream is withdrawn, separately hydrogenated and recycled to the coal extraction reactor as a hydrogen donor solvent. The fractionator bottoms is fed into vacuum distillation unit and then into a Flexicoker to recover additional distillate products. The net liquid products are naphtha (C_5 - 200°C) and fuel oil (200 - 525°C). Propane and butane are coproducts. Sulfur, ammonia and phenols are byproducts.

The residual coke in the Flexicoker is gasified for the production of make-up hydrogen. Additional make-up hydrogen is produced via steam reforming of light hydrocarbon gases (C_1-C_3).

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* Under this configuration, the residence time is 40 minutes in order to obtain higher liquid product yield.

3.4.1 Process Description (cont'd)

For a conceptual plant processing 25,000 t/d of Saskatchewan lignites, the make-up water and electricity requirements from an outside source is estimated at 360 L/s and 110 MW respectively.

In addition to the above process configuration (Design Update Case),ER&E has also examined several other process alternatives. Among them is the so-called "Vacuum Bottoms Recycle Case" with an objective to maximize the mid-distillate and minimize the heavy fuel oil. Under this configuration, a portion of vacuum bottom residues is recycled to the liquefaction reactor which is now operated at higher residence time (60 mins). Because of recycling of vacuum bottoms and higher residence time in the reactor, the equipment cost for coal liquefaction and solvent hydrogenation units also increases significantly (62% when solvent/coal/bottoms ratio of 2.5/1.0/1.0 is maintained in the liquefaction reactor).

3.4.2 Product Yields and Efficiencies

In the EDS process, Texas Big Brown lignite has been extensively tested (10,000 hours of operation) amongst the lignitic coal. Typical product yields of a conceptual EDS plant (Design Update Case) is presented in Table 3-4. Analysis of Texas lignite is given in Appendix B. Both Estevan and Texas Big Brown lignites have nearly similar properties, therefore, product yields and distribution are also similar. Product yields for both coals are 1.8 bbl/t and 2.2 bbl/t on MF and MAF bases respectively. The reported product yields and efficiencies have been adjusted to zero electricity import using a thermal conversion factor of 10,000 kJ/kWh (10,550 kJ/kWh) and a fuel oil equivalent factor of

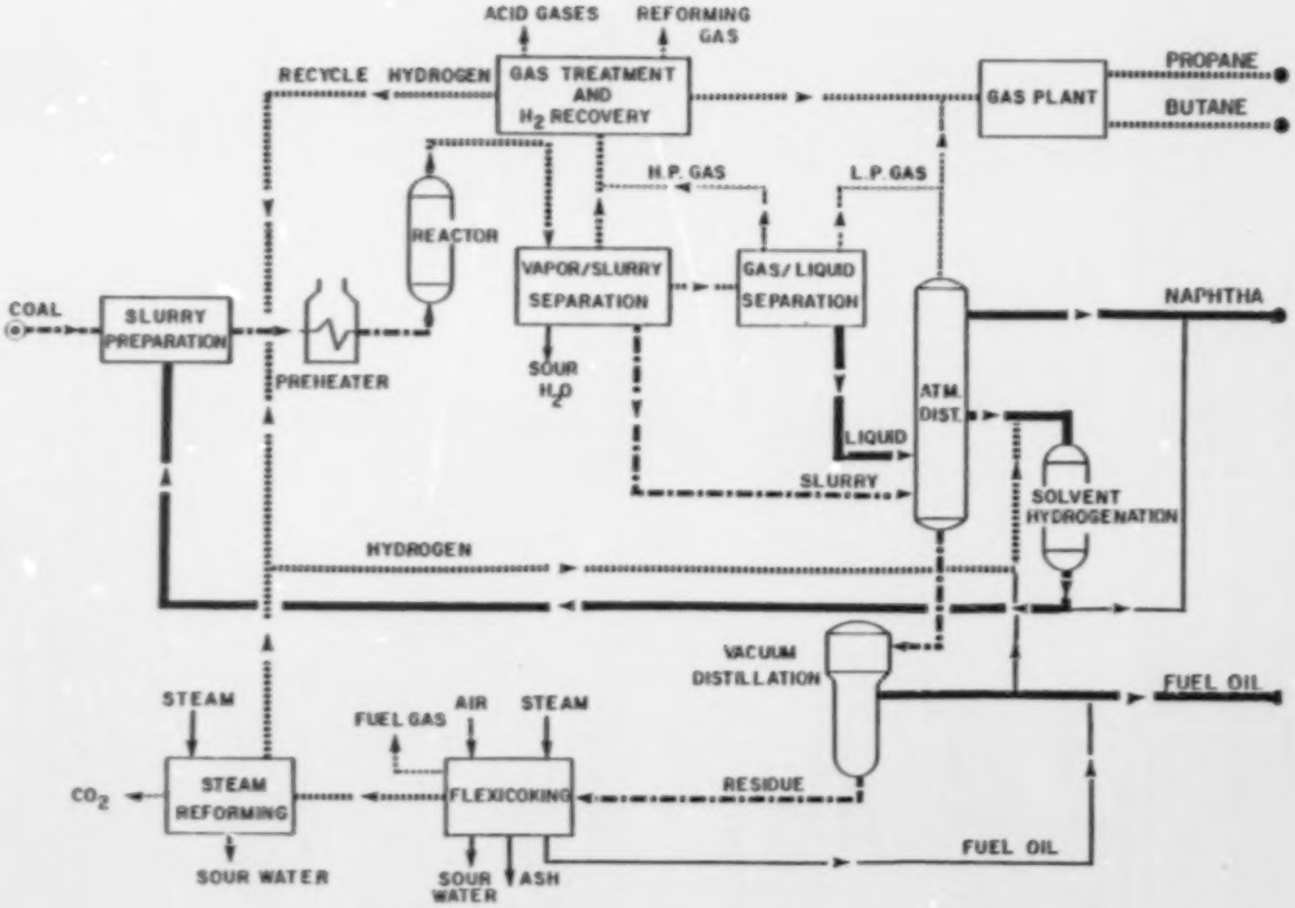

Figure 3-4 Simplified Flow Diagram of the EDS Process

Table 3-4 Product Yield and Efficiencies of the Conceptual EDS Plant

[Basis: 25,000 t/d "as-received" coal]

	Estevan	Texas Lignite Big Brown
Coal (as received)		
Moisture, wt%	35.0	28.4
Ash, wt% of MF coal	15.4	16.4
Coal Fed to Reactor		
Moisture, wt%	4.0	4.0
Ash, wt% of MF coal	15.4	16.4
Input		
MF Coal to Reactor, t/d	15,625	17,212
MF Coal to Gasifiers, t/d	-	-
MF Coal to Offsites, t/d	625	688
Electricity, MW	110	110
Output, t/d (bbl/t)		
SNG	-	-
Propane	123	135
Butane	199	219
Naphtha (C_5-200°C)	2,210 (17,365)	2,435 (19,133)
Fuel Oil (200-525°C)	2,746 (16,782)	3,025 (18,487)
Sulfur	48	200
Phenols	182	53
Ammonia	114	126
Liquid Product Yield (C_5-525°C)*		
bbl/t as-received coal to plant	1.19	1.31
bbl/t MF coal to plant	1.83	1.83
bbl/t MAF coal to plant	2.16	2.16
Naphtha in Total liquid, wt%	44.7	44.7
Product Efficiency*, %		
Overall	53.4	58.3
Liquid	48.2	52.6

* Product yields and efficiencies have been adjusted for zero electricity import using a thermal conversion factor of 10,000 Btu/kWh (10,550 kJ/kWh) and a fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

Source: DOE Report FE-2893-61, Vol.I, March 81; and FE-2893-70, Vol.I, March 31, 1981

3.4.2 **Product Yields and Efficiencies** (cont'd)

6 MM Btu/bbl (6.33 GJ/bbl). Overall product efficiency for Estevan and Texas lignite are 53.4% and 58.3%. Liquid product efficiency are 48.2% and 52.6%.

3.4.3 **Process Status**

The EDS process is being developed as a joint undertaking between the United States Government and a consortium of private industries.

The principal objective of the EDS project is to obtain data to allow a commercial-scale design of the EDS process by 1983. This is being achieved through an integrated process development program involving predevelopment work at bench-scale level, PDU work on a 1-t/d pilot plant and operation of a 227-t/d pilot plant unit, the Exxon Coal Liquefaction Plant (ECLP). The pilot-plant has been operated with Illinois No. 6 coal and Wyoming subbituminous coal, including the "bottoms recycle" mode.

Recently, Texas lignite was also tested but the results are not yet available. According to ER&E, complete analysis of ECLP tests and operational data gathered during its operation will require about six months before final results are obtained and the data released.

The sponsors of the EDS process have decided to terminate the development of the $64-million FLEXICOKING program. Under the new program ER&E is evaluating "vacuum-bottoms" recycling and partial oxidation of vacuum-bottoms in Texaco or Shell gasifiers. The hybrid-boiler concept, in which the bottoms are

3.4.3 **Process Status** (cont'd)

fed to a boiler to generate both steam and to preheat the coal slurry before it is fed to the liquefaction reactor, is also being investigated.

Current projections based on the ECLP operation and the schedule for the alternate processing options indicate that data for a commercial plant design may be available in 1983.

3.4.4 Suitability of Saskatchewan Lignites

Comparing the analyses of Estevan and Texas Big Brown lignites shows that the two lignites are quite similar. Of the ash elements, Estevan contains 15% more SiO_2 but 13% less CaO. Willow Bunch lignite contains about 40% more ash than Estevan lignite (21.0 wt% instead of 15.4 wt% MF coal) and the CaO content is also higher (14.3 wt% instead of 11.7 wt% ash).

In the EDS process, there are two major operational problems associated with the processing of Saskatchewan lignites. The first concern is the vacuum-bottoms viscosity and the second is related to the calcium content of the feed coal which causes reactor wall deposits and line pluggings. High vacuum-bottoms viscosity in the EDS process is related to non-catalytic and milder liquefaction conditions employed in the extraction reactor. The high viscosity of vacuum bottoms due to the nature of the process coupled with the high ash content of the Saskatchewan lignite could create a serious operational problem.

3.4.4 **Suitability of Saskatchewan Lignites** (cont'd)

The second concern is the formation of $CaCO_3$ wall-scales and free-flowing aggregates called "oolites" in the EDS reactor leading to line pluggings and perhaps occasional shut down problems. Operations with Wyodak coal, Texas and North Dakota lignites have shown that the amount of $CaCO_3$ formed is directly related to the initial calcium content present in the feed coal. Since, the CaO content of Saskatchewan lignites is high, processing of the designated lignites would be a problem unless the rate of $CaCO_3$ formation is minimized by pretreating the lignites with SO_2 gas. This aspect has to be experimentally verified for Saskatchewan lignites.

The product yields and efficiency of EDS process (once-through mode) for Saskatchewan lignites are also generally low. Therefore, the liquefaction of Saskatchewan lignites in the EDS process does not appear to be most attractive in comparison to other processes. At the same time, the bottoms-recycle mode suggests that increased light liquid product yield accompanied with reduced bottoms viscosity would improve the overall performance of the EDS process. Therefore, this mode of EDS process should be investigated when the data are available. A discussion held with the process developer is presented in Appendix C-4.

snc

3.5 LIQUID SOLVENT EXTRACTION PROCESS

Two direct catalytic coal liquefaction processes are being developed in England by the National Coal Board (NCB) at the Coal Research Establishment, Stoke Orchard, Cheltenham since 1973; namely the Supercritical Gas Extraction (SGE) and Liquid Solvent Extraction (LSE) processes. Since the SGE process is beyond the scope of this study, only the review of the LSE process is presented.

The LSE process is a two-stage process comprising an "extraction" stage where coal is extracted with a solvent and the extract is separated from mineral matter and ash. In the subsequent "extract hydrogenation" stage, the coal-derived liquids are hydrocracked and stabilized before fractionation. Other main features of the process are:

- The extraction reactor operates at a relatively low temperature and pressure (400°C and 2 MPa), hence capital and maintenance costs are lower.

- Separation of solids (unreacted coal, char, ash) by filtration allows maximum recovery of coal-derived liquid.

- The absence of solids in the coal-derived liquid permits the use of more selective catalysts in the hydrogenation stage, since they are not liable to be deactivated or poisoned by mineral matter.

3.5.1 Process Description

A simplified process flow diagram of LSE process is given in Figure 3-5. Pulverized coal is mixed with recycle distillate oil (oil-to-coal ratio of 2), preheated, and fed into a coal extraction reactor operating at about 400°C and 2 MPa. The

3.5 LIQUID SOLVENT EXTRACTION PROCESS (cont'd)

3.5.1 **Process Description** (cont'd)

reactor effluents are condensed and separated into gases/vapor and solid/liquid slurry streams by conventional flash separation methods. The solid residues (unreacted coal, char, ash) from the slurry are removed by filtration and gasified for the producton of make-up hydrogen for the process. The filtrate (extract) is catalytically hydrocracked and then fractionated to yield naphtha (C_5-200°C), middle oil (200-300°C) and heavy oil (> 300°C). The naphtha and middle oil is further upgraded to gasoline and diesel fuel, and the heavy oil is recycled as a solvent for the process. The coproducts are SNG and LPG. Byproducts are sulfur, ammonia and crude phenols.

Make-up hydrogen is produced via gasification of residual char using Kopper-Totzek* gasifier configuration.

For a conceptual plant processing 25,000 t/d of Saskatchewan lignites, the estimated make-up water requirement is about 500 L/s. Catalyst (Co/Mo) consumption in the hydrocracking unit is about 250,000 kilograms per year. Electricity is generated on site.

3.5.2 **Product Yields and Efficiencies**

Different types and ranks of coals have been tested in the LSE process, however, the main effort has been directed towards the evaluation of high-volatile, non-caking bituminous U.K. coals. Analysis of a U.K. bituminous coal (CRC-702) is given in Appendix B. Some preliminary investigations have been done on coals from other countries such as U.S.A., Australia, Canada and Turkey.

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* In process configuration developed by NCB for U.K. bituminous coal, Shell-Koppers gasifier was considered. For high-ash content Saskatchewan lignites, Koppers-Totzek gasifier has been selected.

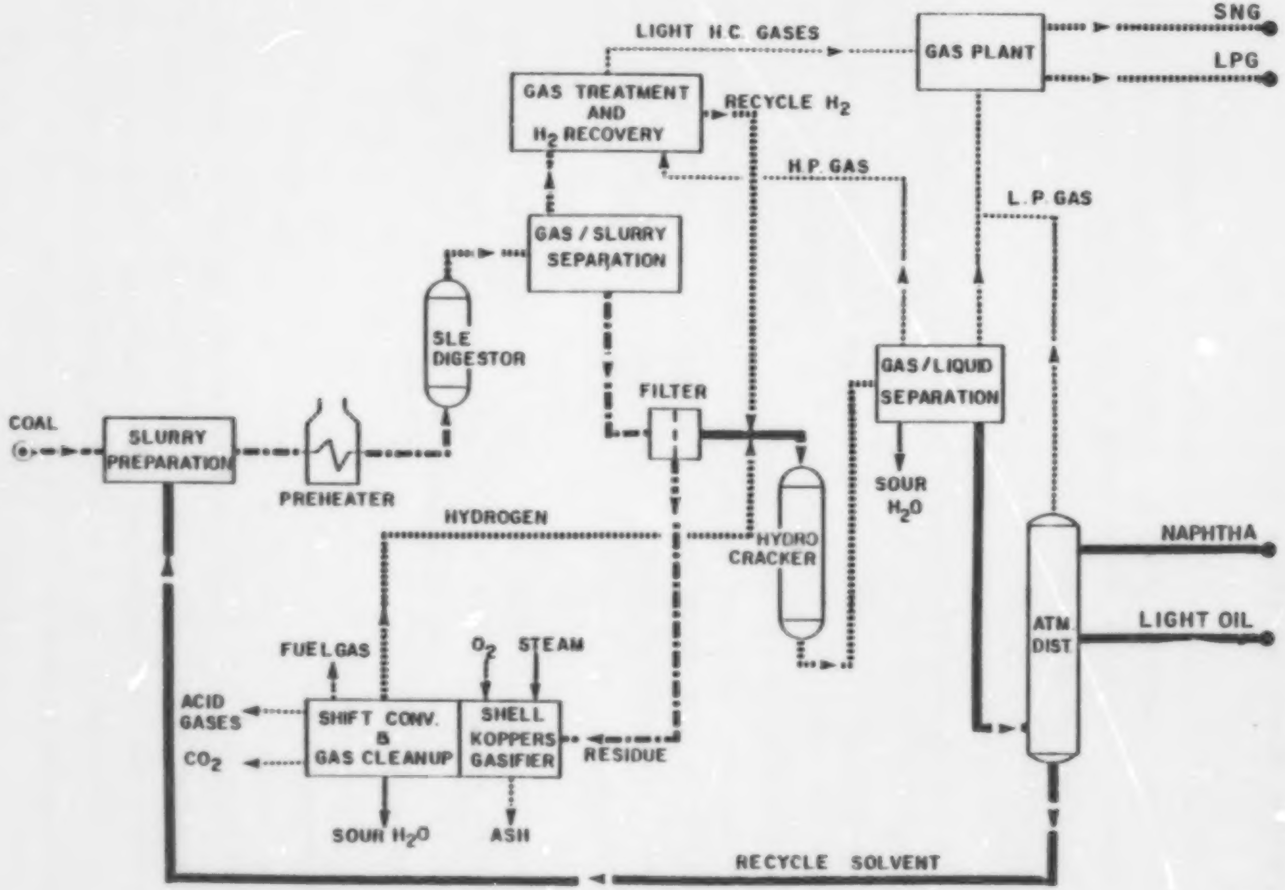

Figure 3-5 Overall Conceptual Configuration of the LSE Process

3.5.2 **Product Yields and Efficiencies** (cont'd)

The product yield and distribution of a self-sufficient, conceptual, commercial LSE plant for U.K. bituminous coal (CRC-702), Estevan and Willow Bunch lignites are presented in Table 3-5. For Saskatchewan lignites, an extraction efficiency of 82% on MAF basis has been assumed.

The results show that the net liquid product (C_5-300°C) yield is about 2.7 bbl/t of as-received U.K. bituminous coal instead of 1.3 bbl/t and 1.1 bbl/t of as-received Estevan and Willow Bunch lignite respectively. (Yields of the Saskatchewan lignites are almost 50% and 60% lower than that of U.K. coal). On an MF basis, the yields are 2.8 bbl/t, 2.0 bbl/t, and 1.7 bbl/t respectively. On an MAF basis, the yields are 3.1 bbl/t of U.K. coal, 2.3 bbl/t of Estevan lignite and 2.1 bbl/t of Willow Bunch lignites.

The "overall" product efficiency for bituminous U.K. coal is 71.0% versus 62% and 61% for Estevan and Willow Bunch lignites. The "liquid" product efficiency are 57.5%, 55.0%, and 54% respectively.

3.5.3 **Process Status**

The National Coal Board has been developing the Liquid Solvent Extraction (LSE) process since 1973. During this period, the process has been tested and demonstrated on a laboratory scale and subsequently in a continuous bench-scale unit having the capacity to process up to 30 kg/h of coal.

Table 3-5 Product Yield and Efficiencies of the Conceptual LSE Plant

[Basis: 25,000 t/d as-received coal]

	U.K. Bituminous (CRC 702)*	Estevan Lignite**	Willow Bunch Lignite**
Coal (as received)			
Moisture, wt%	10.0	35.0	35.0
Ash, wt% MF coal	11.0	15.4	21.0
Coal Fed to Reactor			
Moisture, wt%	5.0	2.0	2.0
Ash, wt% MF coal	11.0	15.4	21.0
Input			
MF Coal to Reactor, t/d	15050	11640	11640
MF Coal to Gasifier, t/d	-	3925	3926
MF Coal to Utilities, t/d	7470	685	684
Electricity, MW	-	-	-
Output, t/d (bbl/d)			
Pipeline Gas (SNG)	1368	-	-
Propane/Butane (LPG)	876	495	447
Naphtha (C_5-200°C)	2969 (22810)	1770 (12230)	1457 (10069)
Mid-Distillate (200-300°C)	5970 (41077)	3049 (19995)	2926 (17161)
Sulfur	114	87	125
Phenol	N/A	N/A	32
Ammonia	245	76	64
Liquid Product Yields (C_5-300°C)			
bbl/t as-received coal to plant	2.7	1.3	1.1
bbl/t MF coal to plant	2.8	2.0	1.7
bbl/t MAF coal to plant	3.1	2.3	2.1
Naphtha in Total Liquid, wt%	33.4	36.7	33.2
Thermal Efficiency, %			
Overall	71.0	62.0	61.2
Liquid Products	57.5	55.0	54.0

Sources:

* NCB Report, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of U.K. National Coal Board", Feb. 1981.

** W.P. Syrett, NCB, Personal Communication to J.A. Mikhlin, SNC Inc., NCB Ref. No. DR 3000, DR 496/2, Oct. 1981; and NCB Ref. No. DR 496/2 January 25, 1982.

3.5.3 **Process Status** (cont'd)

Different techniques of solid/liquid separation such as hydroclones, filtration and solvent precipitation have been evaluated. Of these available methods, the NCB prefers the filtration route because fundamental investigations since 1976 have allowed optimizaton of the extraction reactor operating conditions to give filtration rates that should make this approach viable on a large scale. Moreover, recent developments in hot-filtration technology are leading to new filters, specially designed for coal slurry application. Although these filters are expensive (10% of the total processing cost), the NCB believes that this method of solid/liquid separation is cost effective and has a potential for cost savings. At present, testing has been carried out using 30 kg/h prototype filters.

To demonstrate the process on a larger scale, the decision was made to proceed with a pilot plant (22.7 t/d of coal) for the LSE process. The schedule calls for one-year preconstruction activity, followed by a two-year construction period and a three-year operating and testing period. Process and engineering data for the reliable design and scale-up of a commercial plant will be obtained. The NCB expects the process to reach commercialization readiness stage around 1990.

3.5.4 **Suitability of Saskatchewan Lignites**

The moisture content (35%) of Estevan and Willow Bunch lignites may be intially reduced to 15 wt% by air drying before the coal is fed to a coal/oil slurry preparation unit. Residual moisture present in coal feed is further reduced to less than 2 wt% by slurry-drying method which prevents possible collapsing of porous structure of coal and reduction in coal reactivity.

3.5.4 **Suitability of Saskatchewan Lignites** (cont'd)

In the LSE process, solid/liquid separation is carried out using filtration methods. Therefore, problems associated with the pumping and handling of highly viscous, abrasive, and solids containing residual slurry, as in the case of the vacuum distillation method, are avoided. Instead, the residues are mechanically removed from the filter drums and conveyed to the feed hoppers of the gasifier unit using a screw feeder system. Moreover, the residues contain only about 5 wt% solvent and extract. Consequently the loss of coal-derived liquid is also minimal. Therefore, the net liquid product yields are higher.

The coal-derived liquid obtained in the extraction reactor is filtered for removal of ash and solid residues and is then hydrogenated in a catalytic hydrocracker. This approach has the advantage that for a given coal feedstock it is possible to optimize the operating conditions in each stage so as to produce the desired products with minimum hydrogen consumption. Also, the absence of ash and solid residues in the coal-derived liquid permits the use of more selective catalyst in the hydrogenation stage, since the catalyst is not liable to be deactivated or poisoned by the mineral matter. This aspect is of particular importance in processing of high-ash content coals such as Saskatchewan lignites.

The process developer considers the LSE process suitable for Saskatchewan lignite and an initial set of process data confirming the suitability was submitted to SNC. The NCB response is attached in the confidential Volume V, "Process Developers Data and Information ".

3.6 PROCESS COMPARISON

In the previous section, a brief technical review of five advanced "direct" coal liquefaction processes covering their novel features, coal feed flexibility, conceptual commercial plant configuration, areas of technical concern for processing of Saskatchewan lignites, and process status were described. Product yields and efficiency for a conceptual commercial plant based on process configuration for maximum liquid fuels were also presented. Detailed review and assessment of individual processes, as mentioned earlier, has been presented in Volume II of the study report.

The advanced direct coal liquefaction processes investigated in this study have both specific and common features. Evidently, each process has its novel features and design concept, has differences in operating conditions or level of severity, employs different type of catalysts and process configuration, and hence have different product yields and distribution. In evaluating the advantages of these singular features it is also important to understand the implication of the items on the overall processing, reliability and suitability of the processes for the designated coals. For example, the H-Coal ebullated-bed reactor is considered as a sophisticated system giving better coal conversion and product yield but requiring stringent operational control as compared to upflow tubular reactors. Even in unit operations which are similar in all the processes such as coal/oil slurry preparation, preheating of slurry, high-pressure slurry letdown, flash preparation and liquid/solid separation, the degree of operational complexity is not necessarily the same.

This aspect is related to physical-chemical properties of coal and its liquefaction behaviour in a particular process. Certain criteria such as coal conversion, type of catalyst used and its availability, product yields, and efficiencies which do differentiate the processes can be estimated quantitatively. However, level of operating severity, degree of complexity, etc., which are equally important in ranking of the processes can only be assessed qualitatively unless a designated coal is actually tested in the process. Recognizing the above considerations, a comparative evaluation of the direct coal liquefaction process for designated Saskatchewan lignites is presented below.

3.6.1 Operating Regime

Table 3-6 presents the typical reaction parameters, type of catalysts used, and hydrogen consumption for the direct coal liquefaction processes. All the processes except LSE, employ more or less the same temperature (450-475°C). LSE (extraction stage) operates at a slightly lower temperature (400°C) and considerably lower pressure (2 MPa) compared with the H-Coal and EDS processes (14-15 MPa) or the RAG/Veba and Rheinbraun processes (30 MPa). High pressure is required in the German processes in order to improve the yield of light distillate products and to reduce the preasphaltene content in the recycle heavy oil which would otherwise create serious coking and plugging problems in the preheater tubes.

The reason for such high operating severity required in the RAG/Veba or Rheinbraun processes compared to a similar one-step H-coal process, is because "low quality" disposable red mud (a waste clay obtained from processing of bauxite in aluminum production) is used as a catalyst. Red mud contains only 34% of Fe_2O_3 as catalytically active component, but is cheap and readily

Table 3-6 Typical Reaction Parameters of Advanced Direct Coal Liquefaction Processes

	H-Coal	RAG/Veba	Rheinbraun	EDS	LSE
Temperature, °C	455	475	460	450	400
Pressure, MPa	15.5	30.0	30.0	14.0	2.0*
Catalyst Used					
- Type	Co/Mo (HDS-1442A)	Red Mud (34% Fe_2O_3)	Red Mud (34% Fe_2O_3)	Co/Mo	Co/Mo
- Amount, kg/t MAF coal	0.6	65 (22)	50 (17.5)	N/A#	0.06 ##
Space Velocity, kg/h/m^3	480	530	590	750	N/A
Oil/Coal Ratio	1.2	1.8	1.3	1.6	2.0
Hydrogen Consumption kg/100 kg MAF coal	6.8	6.3	6.0	4.4**	0.0*

# Not available, but the consumption would be in the range of 0.05 to 0.1 kg/t MAF coal.

## Catalyst consumption has been estimated assuming replacement every two years.

* Pressure in subsequent hydrocracking operating is 20 MPa, and H_2 consumption is in the range of 5.0 to 6.5 kg/100 kg MAF coal

** Refers to Design Update case; the total hydrogen consumption including solvent hyrogenation is 5.7 to 6.3 kg/100 kg MAF coal

3.6.1 **Operating Regime** (cont'd)

available. On the other hand, H-Coal uses cobalt molybdate (American Cynamid, HDS-1442A) as a catalyst which is expensive but has better selectivity and hydrocracking activity. However, the "spent" catalyst must be periodically removed from the reactor (normally once a day) and replaced by fresh and/or regenerated catalyst through an on-line catalyst addition and withdrawal system. Results of catalyst regeneration and evaluation tests indicate that at best 80% to 90% of initial catalyst activity may be restored. Catalyst consumption is about 0.5 kg/t MF coal or 0.6 kg/t MAF coal processed and at this rate the annual requirement without regeneration will be approximately 2.6 Gg (2600 t). The supply of catalyst in such an amount is a concern, especially the availability of cobalt.

Production of make-up hydrogen and recovery of unreacted hydrogen represents a significant fraction of both capital and operating costs. Therefore, the amount of hydrogen consumption and extent of hydrogenation achieved is an important consideration. The overall hydrogen consumption for all the processes is nearly the same, about 6.5 kg/100 kg MAF coal. The extent of hydrogenation achieved is related to the product quality which is discussed later, Subsection 3.6.3.

3.6.2 **Product Yield, Efficiency, Utility Requirement**

Table 3-7 presents product yield and distribution of the direct coal liquefaction processes for a 25,000-t/d conceptual plant. The comparison shows that varying amounts of propane/butane, naphtha, mid-distillate and heavy oil are produced. In the RAG/Veba and Rheinbraun processes, no heavy oil is produced as

Table 3-7 Product Flow Rates and Distribution of Advanced Direct Coal Liquefaction Processes
[Basis: 25,000 t/d of as-received lignite to the plant]

	Estevan Lignite					Willow Bunch Lignite		
Processes	H-Coal	RAG/Veba	Rheinbraun	EDS	LSE	H-Coal	LSE*	LSE**
MF Coal, t/d								
- to Reactor	16,250	12,505	10,860	15,625	11,640	14,428	14,152	11,640
- for Fuel Gas and Make-up H_2 Production	-	-	1,555	-	3,925	-	1,267	3,926
- to Utilities	-	3,745	3,835	625	685	1,822	831	684
Electricity from Outside Source, MW	-	160	-	110	[illegible]	-	-	-
Make-Up water, L/s	360	400	525	360	500	360	500	500
Gaseous Products, t/d								
- Propane/Butane	132	952	592	322	495	-	469	447
Liquid Products, t/d								
- Naphtha	2,327	1,058	1,323	2,210	1,770	1,420	1,390	1,457
- Mid-Distillate	2,000	3,553	2,397	2,746	3,049	1,660	2,780	2,926
- Heavy Oil	865	-	-		-	286	-	-
Byproducts, t/d								
- Sulfur	87	57	56	48	87	142	125	125
- Ammonia	145	116	76	114	76	88	64	64
- Phenols	17	-	-	-	-	20	32	32

* Represent product yields for 65% extraction on MAF basis
** Represent product yields for 82% extraction on MAF basis

most of it is recycled to the reactor and the balance is used for diluting the residual slurry to make it pumpable. In the LSE process, all the heavy oil is recycled to the extraction reactor as a solvent. Also, a varying amount of coal is fed to the reactor, to the gasifier for fuel gas and make-up hydrogen production, and to utilities for self-sufficient integrated operation. In the RAG/Veba and EDS processes, electricity is required from outside source.

These variations are due to differences in operating severities, type of catalysts used, amount of heavy oil recycled, process configuration and, of course, due to inherent peculiarities of the process. Mode of operation and method of product recovery (solid/liquid separation) also have a significant impact on the net liquid product yields, particularly for the designated high-ash content Saskatchewan lignites.

For comparison of the processes on a consistent basis, the total liquid product yields and the desired liquid product (naphtha and mid-distillate) yield expressed in unit coal feed basis, i.e., bbl/t coal fed to the reactor and bbl/t coal supplied to the plant, are presented in Table 3-8. Product yield per tonne of coal supplied to the plant has been adjusted to zero electricity input for the RAG/Veba and EDS processes using the conversion factor of 10,000 Btu/kWh (10,550 kJ/kWh) and a fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

For both Estevan and Willow Bunch lignites, the results show that the LSE process has the highest product yield per tonne of coal fed to the reactor followed by Rheinbraun, RAG/Veba, H-Coal and

Table 3-8 Product Yield and Efficiency of Advanced Direct Coal Liquefaction Processes

Coal	Estevan Lignite					Willow Bunch Lignite		
Process	H-Coal	RAG/Veba	Rheinbraun	EDS	LSE**	H-Coal	LSE*	LSE**
Percentage of Total Coal Fed to Reactor	100	77	67	96	72	89	87	72
Total Liquid Product Yield								
bbl/t as-received coal to reactor	1.52	1.57	1.64	1.42	1.79	1.11	1.19	1.51
bbl/t MF coal to reactor	2.34	2.41	2.52	2.19	2.75	1.71	1.83	2.33
bbl/t MAF coal to reactor	2.76	2.85	2.98	2.59	3.25	2.17	2.32	2.95
Adjusted Total Liquid Product Yield #								
bbl/t as-received coal to plant	1.52	0.95	1.10	1.19	1.29	0.99	1.03	1.09
bbl/t MF coal to plant	2.34	1.47	1.69	1.83	1.98	1.52	1.59	1.68
bbl/t MAF coal to plant	2.76	1.73	2.00	2.16	2.34	1.93	2.02	2.12
Adjusted Total Liquid Product Yield Including LPG ##								
bbl/t as-received coal to plant	1.56	1.25	1.28	1.29	1.43	0.99	1.17	1.22
bbl/t MF coal to plant	2.40	1.92	1.97	1.99	2.20	1.52	1.80	1.87
bbl/t MAF coal to plant	2.84	2.27	2.33	2.35	2.60	1.93	2.28	2.37
Adjusted Desired Liquid Product Yield (Naphtha and Mid-Distillate)								
bbl/t as received coal to plant	1.29	0.95	1.10	1.03	1.29	0.92	1.03	1.09
bbl/t MF coal to plant	1.99	1.47	1.69	1.58	1.98	1.41	1.59	1.68
bbl/t MAF coal to plant	2.36	1.73	2.00	1.87	2.34	1.79	2.02	2.12
Product Efficiency								
Overall	62.0	55.7	52.2	53.4	62.0	46.8	58.9	61.2
Total Liquid Product (excluding LPG)	59.0	44.0	43.2	48.2	55.0	45.7	51.4	54.0
Desired Liquid Product	49.7	44.0	43.2	35.4	55.0	45.7	51.4	54.0

# For RAG/Veba and EDS process, the yields in bbl/t coal to plant has been adjusted to zero electricity import using 10,000 Btu/kWh (10,550 kJ/kWh) and fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

## For LPG produced in the processes, the amount of barrels have been added using fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

* Represent product yields for 65% extraction on MAF basis.
** Represent product yields for 82% extraction on MAF basis.

3.6.2 **Product Yield, Efficiency, Utility Requirement** (cont'd)

EDS processes. Higher liquid yields on a reactor basis for the LSE process are primarily due to maximum recovery of coal-derived liquids in the solid/liquid separation step which is carried out by filtration as opposed to flash separation and vacuum distillation methods used in the other processes. The residual char (or solid residues) in the LSE process contains only 5 wt% liquids compared to at least 50 wt% liquids required in other processes in order to pump the vacuum bottoms to the gasifiers for make-up hydrogen production.

On an overall plant basis, the total liquid product yields including LPG are comparable for the LSE and H-Coal processes in the case of Estevan lignite but is higher for LSE when processing Willow Bunch lignite. The desired liquid product yields (naphtha and mid-distillate*) are also highest in the LSE process for Willow Bunch lignite, and approximately the same in LSE and H-Coal processes for Estevan lignite. Plant efficiency of desired liquid products for both Willow Bunch and Estevan lignites is higher for the LSE process. The German processes (Rheinbraun and RAG/Veba) and the EDS process have lower yields and efficiencies.

3.6.3 **Product Quality**

Data on quality of products obtained from processing of lignites are not available since most of the process developers have directed their efforts towards testing of bituminous and subbituminous coals. Therefore, for comparative purposes, the data on product quality available for bituminous coals are

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* Naphtha and mid-distillate are considered as desired products for upgrading transportation fuels.

3.6.3 **Product Quality** (cont'd)

presented in Table 3-9. A typical analysis of the feed coals which to a certain extent dictate the final quality of the products is given in Appendix B.

The results show that H-Coal and LSE processes produce better quality products. The LSE products also have the least amount of heteroatoms, the mid-distillate is in the diesel range (200-300°C) and no heavy oil is produced, so relatively less severe upgrading would be subsequently required. This is because the two-stage approach of LSE process allows more selective operating conditions to be used in the catalytic hydrocracking step, and in fact, partial upgrading is also achieved.

The EDS process produces naphtha (C_5-200°C) and heavy fuel oil (200-525°C). The products have somewhat lower hydrogen content and higher heteroatom content. Naphtha is of good quality but the fuel oil is heavy and has poor quality.

The RAG/Veba products are relatively heavier, have even lower hydrogen content and very high heteroatom content, particularly the amount of oxygen which is present mainly as undesirable phenols. Sulfur content appears to be lower than in H-Coal or EDS, but this is due to a lower amount of sulfur present in Ruhr coal (1.6 wt% S) compared with Illinois No. 6 coal (3.5 wt% S). The products are heavier and poorer in quality despite severe operating conditions. The hydrogen consumption is as high as other processes (see Table 3-6) indicating that expensive hydrogen is somewhat wasted and that the red mud catalyst is not very effective in hydroconversion, hydrodesulfurization, denitrogenation and deoxygenation reactions in the liquefaction reactor.

Table 3-9 Product Quality for Advanced Direct Coal Liquefaction Processes

Process (Coal)	H-Coal[1] (Illinois No.6)	RAG/Veba[2] (Ruhr Coal)	Rheinbraun (Rhenish Brown)	EDS[3] (Illinois No.6)	LSE[4] (CRC-702)
Naphtha					
API Gravity	52.3	29.2		52.3	41.0
Elemental Analysis, wt%					
C	85.34	85.05	N/A	85.2	87.0
H	13.79	10.80		13.16	12.4
O	0.56	3.60		1.15	0.5
N	0.24	0.24		0.43	0.2
S	0.07	0.03		0.06	0.001
Mid-Distillate					
API Gravity	18.5	11.2			23.0
Elemental Analysis, wt%					
C	88.35	88.15	N/A	*	88.7
H	10.41	8.80			10.6
O	0.99	2.60			0.5
N	0.47	0.85			0.2
S	0.08	0.08			0.001
Heavy Oil					
API Gravity	4.9	2.00		5.9	**
Elemental Analysis, wt%					
C	89.42	88.15	N/A	87.56	
H	8.56	7.30		8.83	**
O	1.31	2.20		2.30	
N	0.63	1.45		0.54	
S	0.08	0.39		0.77	

* In EDS process, naphtha (C_5-200°C) and heavy fuel oil (200-525°C) is produced .

** In LSE process, the mid-distillate in the temperature range of 200-300°C and no heavy oil is produced as all of it is recycled to the extraction reactor

Sources:

1. EPRI Report AF-1297, December 1979.
2. G. Escher et al., "Ruhrkohle/Veba Oel Technology for Hydrogenation of Hard Coal", paper presented at Annual Fall Conference of CSChE, Edmonton, Oct. 19-22, 1980
3. DOE Report No. FE-2893-61, Vol.I, March 1981
4. NCB Report to SNC Inc. "Coal Liquefaction for Transport Fuels using the Liquid Solvent Extraction of the UK National Coal Board", Feb. 1981.

3.6.4 Coal Feed Flexibility Process Complexity and Reliability

All the processes have been tested under a wide range of operating conditions but with different types of coals and at different scales of operation such as bench scale, PDU, and pilot plant.

Table 3-10 presents the type and rank of coals investigated in the various processes. The analyses of these coals are attached in Appendix B. Table 3-11 gives the extent of testing and demonstration level for the low-rank coals. The results show that in so far as low-rank coals are concerned, all the processes except EDS have only been tested and demonstrated on a bench scale. Therefore, coal feed flexibility of these processes for commercial-scale operation with Saskatchewan lignites cannot be predicted in realistic terms. The EDS process is the only process which has tested both Texas and North Dakota lignites in 1-t/d unit and recently Texas lignite was also tested in 227-t/d pilot plant. The results of the pilot plant runs are not yet available but successful operations were carried out.

In terms of mechanical reliability of the processes, a list of critical operating areas pertaining to the processing of Saskatchewan lignites is given in Table 3-12. In most cases, the problem arises from high severity of operating conditions coupled with physical/chemical properties of the Saskatchewan lignites. In some cases, further development is required for improved and reliable operation, although the extent of development would vary from process to process. A number of engineering approaches are available to deal with each problem area, but cost not the probability of technical success is the issue. From Table 3-12, it

Table 3-10 Types of Coals* Tested in Advanced Direct Coal Liquefaction Processes

	H-Coal	RAG/Veba	Rheinbraun	EDS	LSE
Bituminous	Illinois No.6 Kentucky No.9	Ruhr Coal	-	Illinois No.6	U.K. Coal (CRC-702)
Subbituminous	Wyodak	-	-	Wyodak Wandoan	
Lignite/Brown	Texas North Dakota Australian Brown		Rhenish Brown	Texas North Dakota	Australian Brown

* Coals that were tested briefly or only in preliminary tests are not included. A complete list of coals tested is covered in the detailed technical review given in Volume II of the report.

Table 3-11 Level of Testing and Demonstration for Low-Rank Coals

	H-Coal	RAG/Veba	Rheinbraun	EDS	LSE
Texas Lignite	Bench-Scale (11 kg/h) 705 hours	-	-	Bench-Scale (2 kg/h) PDU (1 t/d) Pilot Plant (227 t/d)	
North Dakota Lignite	Bench-Scale (11 kg/h) 505 hours	-	-	Bench-Scale (2 kg/h) PDU (1 t/d)	
Rhein Brown	-	-	Bench-Scale (11 kg/h)	-	-
Australian * Brown	Microautoclave, Bench-Scale (11 kg/h)	-	-	-	Bench-Scale (30 kg/h)

* In H-Coal process, Muswellbrook coal (Australian brown) has been tested in microautoclave unit, whereas brown coal from Gelliondale field has been tested in bench-scale unit.

Table 3-12 List of Critical Coal Operating Areas in Direct Coal Liquefaction Processes

	H-Coal	RAG/Veba	Rheinbraun	EDS	LSE
Coal Drying and Moisture Removal	X	X		X	X
Slurry Feed Pumps and Pressurizing	X	XX	XX	X	
Slurry Preheating	X	XX	XX	XX	X
Reactor Control	XX	X	X	X	X
Heat Recovery from Reactor Effluents	X	XX	XX	X	
Pressure Let-Down Valves and Flash Separation of Slurry	XX	XX	XX	X	
Catalyst Deactivation and Generation	X				
Solid/Liquid Separation	X	X	X	X	XX
Pumping of Vaccuum Bottoms (Residues) and Removal of Solids in LSE process	X	XX	XX	X	X
Gasification of Residual Slurry (or Residues)	X	XX	XX	X	X
Material Erosion and Corrosion	X	XX	XX	X	X
Environmental Impact, Health and Safety	X	X	X	X	X

Blank = Not considered as critical
X = Less critical
XX = Critical

3.6.4 Coal Feed Flexibility Process Complexity and Reliability (cont'd)

appears that LSE process has the least critical areas and as such offers maximum reliability for processing of Saskatchewan lignites. The RAG/Veba and Rheinbraun processes employ severe operating conditions and use red mud catalyst which together with high-ash content of Saskatchewan lignites is likely to affect the operability of equipment items in several areas.

3.6.5 Status of Process Development

H-Coal, EDS and RAG/Veba process developers have advanced their processes to a pilot-plant scale (225-t/d H-Coal plant at Catlettsburg, 227-t/d EDS plant at Baytown, 200-t/d RAG/Veba plant at Bottrop).

The LSE process has been demonstrated on a continuous bench-scale installation having a capacity of processing 0.7 t/d (30 kg/h) of coal. Some of the process steps such as 2 t/d solvent extraction of coal and 0.7 t/d filtration unit have also been separately demonstrated on other installations. To demonstrate the process on a larger scale, a decision was made to proceed with the pilot plant (22.7 t/d of coal) for the LSE process. The schedule calls for one-year preconstruction acitivity, followed by a two-year construction period and a three-year operating and testing period.

Rheinbraun process has been tested on a bench-scale unit of 0.25 t/d of coal only. Process design of a pilot plant (360 t/d of coal) is underway.

3.6.5 **Status of Process Development** (cont'd)

All the process developers (except Rheinbraun) offer commercialization by 1990. Rheinbraun is scheduled for commercial readiness by 1995.

Unfortunately, the current economic difficulties in the developed countries, including U.S.A., Federal Republic of Germany and U.K., coupled with the temporary world oil glut have slowed down the development of all the processes considered on the Study and may delay their commercialization target.

At present, the U.S. Department of Energy is supporting the experimental program at Catlettsburg H-Coal plant. The program is to be completed at the end of 1982 and the process developer is looking for other sponsors to continue the operation of the plant. In case further funding will not be available, the plant will be dismantled.

At the EDS pilot plant in Baytown, the experimental program is also completed and as it was announced recently the plant will be on shut down shortly and probably dismantled.

The NCB, process developer of the LSE process, is looking for new sponsors for funding of the LSE pilot-plant project, since British Petroleum, one of the sponsors of LSE process announced recently that it will not be able to support the project further in view of financial constraints.

The RAG/Veba pilot plant at Bottrop is in operation. However, as announced recently by the process developer, funding for further development of this process is in jeopardy.

3.7 PROCESS RANKING

Ranking of "direct" advanced coal liquefaction processes is made based on process comparison presented earlier and in accordance with the selection criteria set out by the Study Committee. The objective is to select a direct coal liquefaction process for further techno-economic evaluation (Stage II of the Study).

The LSE process appears to have the highest priority, especially for Willow Bunch lignite which is designated by the Study Committee as a feedstock for techno-economic assessment of preselected coal liquefaction processes (Stage II of the Study). The main concern in this respect is the development status of the LSE process particularly due to the recent financial difficulties experienced by the process developer for continuation of the pilot-plant program.

The product yield, overall and liquid product efficiencies for the LSE process are higher than the next competing H-Coal process even if a modest 65% extraction on MAF basis is considered. The quality of primary products in the LSE process is also better and would require less upgrading to transportation fuels. The LSE process employs milder operating conditions during the extraction phase. Rapid deactivation of the catalyst in the H-Coal process due to high ash content in Saskatchewan lignites could affect the catalyst regeneration efficiency and may lead to higher catalyst consumption.

The results presented earlier in Table 3-8 show the effect of high-ash content on product yields and efficiencies of the processes employing flash separation and distillation methods for solid/liquid separation. Reduction of ash content to a desirable level by pretreatment/beneficiation could possibly improve the product yields and change the relative merit of the processes. This aspect is beyond the scope of the present study and must be experimentally verified, along with cost/benefit analysis to determine if it is feasible to carry out such pretreatment for Saskatchewan lignites.

The third candidate appears to be the EDS process (once-through mode). It has lower product yields and efficiencies than LSE or H-Coal but has higher liquid product yields and efficiencies than the RAG/Veba or the Rheinbraun process. The EDS naphtha is of better quality but the fuel oil is inferior to the RAG/Veba product. EDS has also tested high-ash content Texas lignite in its pilot plant.

The next candidate is RAG/Veba followed by the Rheinbraun process. Both processes are a modified version of classical I.G. Farben process and are inherently similar except that the RAG/Veba process has been developed for low-ash content bituminous Ruhr coal and Rheinbraun process for high-moisture and low-ash content Rhenish brown coal. The RAG/Veba process has reached pilot-plant stage (225 t/d of coal) whereas the Rheinbraun process is on a bench scale. Both processes employ severe operating conditions and have higher hydrogen consumption and yet the quality of product is low. Liquid product yields and efficiencies for Saskatchewan lignites is also very low. The response of the respective process developers of the RAG/Veba and Rheinbraun processes confirming the unsuitability of the process for as-received Saskatchewan lignites is attached in Appendix C.

INDIRECT COAL LIQUEFACTION

4.0 OVERVIEW OF ADVANCED OR COMMERCIALLY PROVEN PROCESSES AND THEIR SUITABILITY FOR DESIGNATED SASKATCHEWAN LIGNITES

4.0 INDIRECT COAL LIQUEFACTION: OVERVIEW OF ADVANCED OR COMMERCIALLY PROVEN PROCESSES

"Indirect" coal liquefaction, as distinguished from "direct" coal liquefaction (hydrogenation) involves gasification of the coal to produce synthesis gas ($CO+H_2$) as the first step followed by conversion of the purified syngas to liquid hydrocarbons or to methanol in the subsequent steps. At present, there are three main process alternatives for the conversion of coal to liquid fuels via indirect coal liquefaction: Fischer-Tropsch synthesis, methanol synthesis and Mobil Methanol-to-Gasoline process. The following process alternatives in indirect coal liquefaction were selected by the Study Committee:

- SASOL-type plant with Fischer-Tropsch synthesis using Synthol reactor (F-T process).
- Coal-based Methanol-to-Gasoline plant with Mobil MTG process using fixed-bed reactor.
- Coal-based plant for production of pure methanol (99.9% purity) as tranportation fuel and/or blendstock for gasoline.

In Figure 4-1, a simplified diagram for the above three routes are presented. A detailed description of these three process alternatives in indirect coal liquefaction are presented in Volume III, "Assessment of Indirect Coal Liquefaction Processes Considered for Commercialization by 1990".

Coal-based plant with Fischer-Tropsch synthesis using Synthol reactor is a commercially proven technology. Three coal-based plants: SASOL-I, (10,000 to 15,000 t/d of coal), SASOL-II and SASOL-III plants (design capacity 40,000 t/d of coal each) are operating in South Africa.

Coal-based methanol production is also proven on a commercial scale. A 3000 t/d of coal plant producing methanol and ammonia is in operation since 1974 (AECI plant, Modderfontein, South Africa).

The Mobil MTG process is being developed in two reactor modes: fixed-bed and fluidized-bed. The fixed-bed reactor mode selected for the current study has been demonstrated in a small pilot plant producing 4 bbl/d of gasoline. However, the process developer, Mobil Research and Development Corporation, U.S.A., considers that the process demonstration program is complete and the process is ready for scale-up to a commercial size. At present, based on the experimental results obtained in the 4 bbl/d pilot plant, a detailed engineering design of a commercial plant producing 14,000 bbl/d of gasoline to be built in New Zealand is completed. Natural gas will be used as a feedstock for the New Zealand commercial plant. The construction of the plant is scheduled for completion in 1986.

The fluidized-bed reactor mode of the Mobil-MTG process was also demonstrated on a 4 bbl/d of gasoline scale. Due to the difficulty of scaling-up this type of reactor, a 100 bbl/d plant is being constructed at present in the Federal Republic of Germany for further demonstration of the fluidized-bed mode of

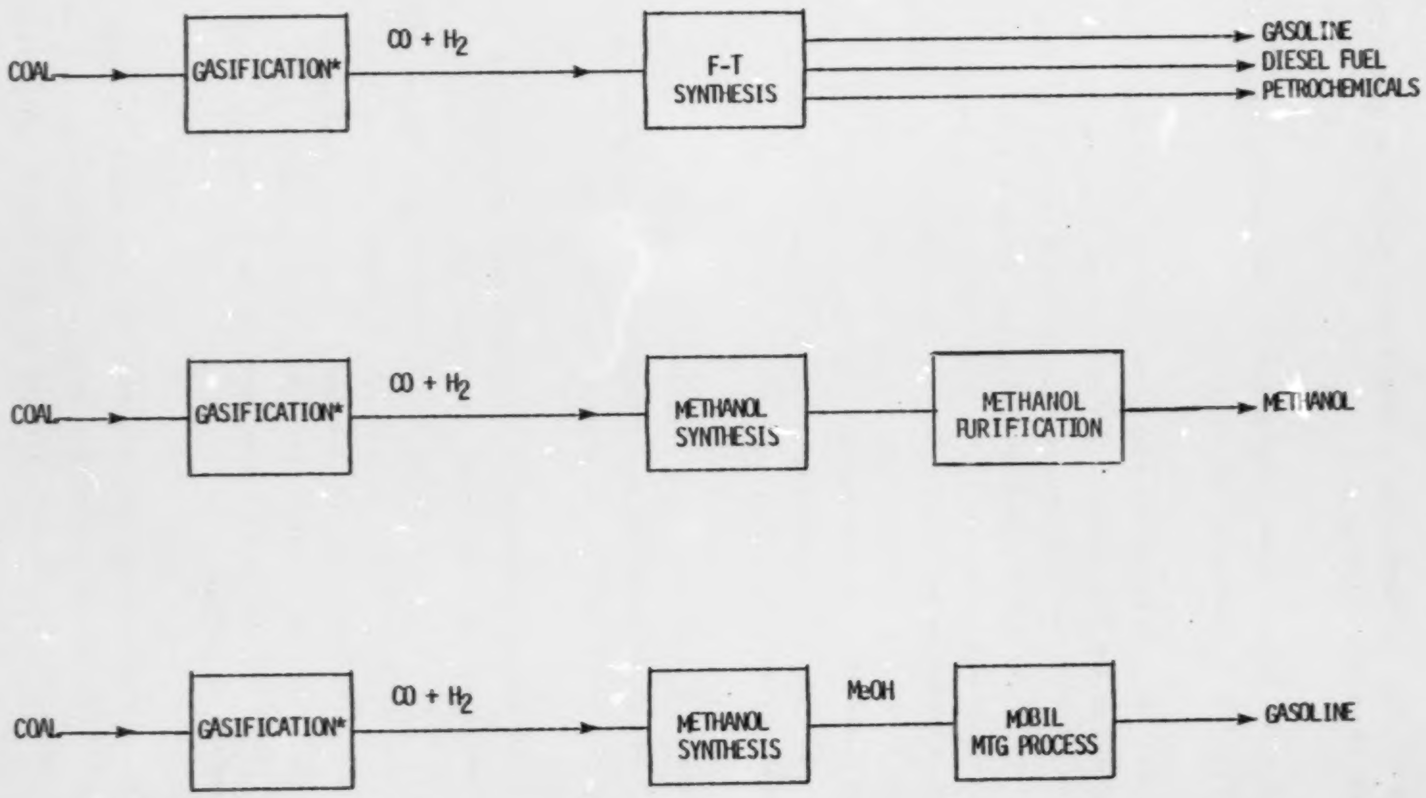

Figure 4-1: Simplified Diagrams of Indirect Coal Liquefaction Alternatives via Fischer-Tropsch (F-T) Synthesis, Methanol Synthesis, and Mobil-MTG Process.

* coal gasification block include gas treatment, shift conversion, etc.

the process before commercialization. The completion of plant construction is scheduled for 1983.

Analyzing the schematics in Figure 4-1, it is clear that in order to determine the suitability of Saskatchewan lignites for liquefaction through an indirect route, emphasis should be made on selection of an appropriate gasification process (including raw gas pretreatment required). After coal is converted into synthesis gas, any of the three process alternatives considered for the study may be technically feasible and the final overall process selection should be made on the basis of the product yield and quality required, overall process economics, and readiness toward commercialization.

As stated in the selection criteria, a single gasification process had to be selected for techno-economic evaluation of all three process alternatives considered in this study

The six commercially proven or advanced coal gasification processes selected by the Study Committee for prescreening are listed in Table 4-1. All three major types of gasifier configuration used at present for production of syngas from coal, i.e., fixed bed, fluidized bed and entrained flow, are considered in the study. Also, all gasification processes listed in Table 4-1 are considered in an oxygen-blown mode which is usually applied to production of syngas.

The Lurgi, Koppers-Totzek and standard Winkler gasifiers have a history of reliable operation on a large commercial scale and are usually referred to in the literature as "first generation"

Table 4-1 Advanced Coal Gasification Processes for Production of Synthesis Gas

Gasifier	Type	Process Developer
Lugi (L)	Fixed Bed (Dry Bottom)	Lurgi Kohle and Mineralöltechnik, F.R.G.
British Gas/Lurgi (BGC/L)	Fixed Bed (Slagging)	British Gas Corporation U.K.; and Lurgi, F.R.G.
Koppers-Totzek (GKT)	Entrained Flow	Krupp-Koppers, F.R.G.
Texaco	Entrained Flow	Texaco Developement Corporation, U.S.A.
Winkler (HTW)	Fluidized Bed	Rheinische Braunkohlenwerke AG (Rheinbraun), F.R.G.*
Westinghouse (W)	Fluidized Bed	Westinghouse Electric Corporation, U.S.A.

* Developer of high pressure and high temperature Winkler gasification process (HTW process); the derivation of the standard Winkler coal gasification.

gasifiers. The term "second generation" gasifiers is used to describe either derivations of the original first generation designs (e.g., BGC/L and HTW gasifiers) or new concepts (e.g., Texaco, Westinghouse gasifiers).

Detailed description and status of gasifiers listed in Table 4-1 are given in Volume III, Chapter 1 of the study report.

4.1 FACTORS AFFECTING SYNGAS PRODUCTION FROM COAL AND SUITABILITY OF SASKATCHEWAN LIGNITES FOR SELECTED PROCESSES

Coal properties and several other factors are affecting gasifier performance and selection of an appropriate coal gasification process.

Ash content in the coal and ash composition are among the coal properties of main concern. Also, feed coal size distribution, oxygen content, reactivity and other coal properties have an influence on gasifier performance.

Among the other factors which may affect gasifier selection for syngas production are raw gas composition (especially methane content, hydrogen-to-carbon monoxide ratio, sulfur content) operating pressure, oxygen consumption and oxygen-to-steam ratio. Also, process stability, raw gas output per coal gasification unit, etc., should be taken into account for a commercial plant setting.

4.1.1 Feed Coal Size and Size Distribution

Limitations on fines in the coal feed relate mostly to fixed-bed gasifiers, i.e, Lurgi dry-bottom and British Gas/Lurgi gasifiers, although the BGC/L has made advances in this area. The lump size coal is fed to a fixed-bed gasifier via a lock-hopper system at the top of the vessel. In the Lurgi gasifier, coal grading is required to limit the quantity of fines in the coal and to maintain a ratio of not more than 10:1 of largest to smallest size. Although some fines can be tolerated, this should not exceed 7%, as was demonstrated by the process developer in the gasification test at Westfield, Scotland. If the amount of fines

4.1.1 **Feed Coal Size and Size Distribution** (cont'd)

is higher, as for designated Saskatchewan lignites (see Appendix A), the fines must be separated. The possibility of fines briquetting before gasification may be considered. The cost of briquetting of coal fines were estimated to be about $1.93/tonne of as-received coal (mid-1979 dollars)*.

The fluidized-bed and entrained-flow gasifiers will usually tolerate fines in the coal feed since the coal has to be ground or pulverized in this case before feeding into the gasifier (for details, see Volume III, Chapter 1).

4.1.2 **Moisture Content in the Feed Coal**

Gasifier performance is penalized by high moisture content in the feed coal. For gasifiers equipped with dry coal feeding systems, the level of coal predrying is usually dictated by overall process economics. For a wet coal feeding system as used in the Texaco gasifier, the level of achievable concentration of coal in the feeding system and the influence on overall process economics must also be taken into account.

Coal predrying to 15 wt% moisture was selected as economically reasonable by Lurgi for gasification of a high moisture lignite (e.g., 40 wt% moisture in as-received Shaunovan coal, Saskatchewan, Canada), although Lurgi gasification test with as-received North Dakota lignite (36.0 wt% moisture) resulted in a gas outlet temperature above the dew point of about 65°C*.

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* Saskmont Eng., "Shaunavon Coal Utilization Study - Phase II", Volume II, SPC Report, July 1980.

4.1 FACTORS AFFECTING SYNGAS PRODUCTION FROM COAL AND SUITABILITY OF SASKATCHEWAN LIGNITES FOR SELECTED PROCESSES (cont'd)

4.1.2 **Moisture Content in the Feed Coal** (cont'd)

Coal predrying to 15 wt% moisture was also considered as economically reasonabale for processing lignite in the Westinghouse gasifier*.

The usual level of coal predrying selected by the HTW process developer for gasification of high moisture Rhenish brown coal (60 wt% moisture on as-received basis) is between 12 to 16 wt%.

Predrying of lignite to 8 wt% moisture was selected by Krupp-Koppers for the commercial operation of GKT gasifiers in Greece (coal-based ammonia plant using Ptolomais lignite as the feedstock).

Coal predrying to 15 wt% moisture may also be considered for the designated Saskatchewan lignites having 35 wt% moisture content on as-received basis.

In the Texaco coal gasification process, a concentrated slurry of coal particles and water is fed into the upper part of the gasifier. Usually, coal concentration in the slurry is at about 60 to 65 wt% (MF basis). In the tests with North Dakota lignite, coal concentration in water slurry of 55 wt% (MF basis) has been reported by the process developers, without predrying of coal. However, coal tests on grinding and slurrying in water of Willow Bunch lignite carried out by Texaco revealed that this lignite has to be completely dried before slurrying in water in order to achieve

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* Direct Communication by J.A. Mikhlin, SNC Inc. with process developer, January 1982.

4.1.2 Moisture Content in the Feed Coal (cont'd)

the coal concentration in the slurry of 48 to 49 wt% (MF basis), see Appendix C-5*. The required deep drying of Willow Bunch lignite may damage its porous structure and result in corresponding decrease in coal reactivity, in addition to a substantial extra processing cost (for details, see Volume III, Chapter 1).

4.1.3 Ash Content

High ash content of Estevan lignite (15.4 wt% MF coal), and Willow Bunch (21.0 wt% MF coal) will result in decreased efficiency of the gasification process and higher oxygen consumption to keep the required operating temperature. The effect of high ash content will be more pronounced, the higher the operating temperature (see Table 4-2).

Trouble-free operation with high-ash-content lignite was commercially proven in Lurgi dry-bottom gasifiers (up to 26.6 wt% ash MF coal), as well as in the standard Winkler gasifiers (not considered in the present study for prescreening because of relatively low coal throughput and carbon conversion obtained in this first generation gasifier). Brown coal with 3 wt% ash (MF basis) and lignite with 7 wt% ash (MF basis) have been processed in HTW and Westinghouse processes on a pilot plant level (32 t/d coal for HTW and 24 t/d coal for Westinghouse).

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* The lowest coal concentration in the feed slurry of Willow Bunch lignite to achieve still a reasonable economics of Texaco coal gasification process, as considered by the process developer.

4.1.3 Ash Content (cont'd)

The fusion or melting point of coal ash is another important property which may affect gasifier performance. Coals with high ash content and having a high melting point of ash could show poor economics. Coals with very high ash-fusion temperature could be difficult to slag even in entrained-flow gasifiers. On the other side, low ash-fusion temperature would require increasing steam consumption for such as Lurgi dry-bottom gasifier to keep the ash temperature below melting point.

The ash softening temperature in Estevan and Willow Bunch lignites is around 1100 to 1200°C and should not cause problems in normal performance of any gasifiers considered for the study.

Ash composition may also sufficiently affect gasifier performance, especially if ash contains sodium and potassium acids which in many cases appear to have a direct relationship to the amount of fouling that could be expected during lignite gasification. In the designated Saskatchewan lignites, the Na_2O and K_2O contents in the ash are as follows.

	Estevan	Willow Bunch
wt% of Ash		
Na_2O	5.0	1.8
K_2O	0.5	1.3
wt% of MF coal		
Na_2O	0.77	0.38
K_2O	0.08	0.27

4.1 FACTORS AFFECTING SYNGAS PRODUCTION FROM COAL AND SUITABILITY OF SASKATCHEWAN LIGNITES FOR SELECTED PROCESSES (cont'd)

4.1.3 **Ash Content** (cont'd)

These amounts of N_2O and K_2O in the ash may cause problems in gasifier performance. However, it is almost impossible to predict the behaviour of ash and/or it's components under gasifier conditions based on analytical estimation of ash only. Therefore, unless the gasification process has been demonstrated on a commercial scale processing similar coals, the process developers require coal tests under simulated or real gasification conditions to predict the coal and ash behaviour in a gasification process.

4.1.4 **Coal Reactivity**

Lignites are usually characterized as most reactive coals as compared with higher rank coals (bituminous and subbituminous). However, as mentioned earlier, excessive pretreatment of lignite, e.g., hot-air deep drying, high-temperature thermal drying may result in decreased reactivity of coal during gasification.

4.1.5 **Raw Gas Composition**

4.1.5.1 Methane Content

Methane content in the raw gas has to be at the minimum possible level if coal gasification is aimed for production of synthesis gas. This is because each mole of methane produced represents a loss of 3 moles of syngas.

4.1.5.1 Methane Content (cont'd)

Typical compositions of raw gas produced in oxygen-blown gasifiers selected for prescreening in this study, along with operating parameters (pressure, temperature, oxygen and steam consumption) are presented in Table 4-2. Raw gas composition of all the gasifiers (except for BGC/L) are for lignites. BGC/L data are for bituminous Pittsburgh No.8 coal since lignites have not been tested in this gasifier. Raw gas composition of the HTW gasifier has not been reported by the process developers (except for the methane content obtained by processing Rhenish brown coal). For comparison of operating parameters (mostly steam and oxygen consumption) for BGC/L and Lurgi dry bottom gasifiers, the parameters for these two alternatives of fixed-bed gasifiers are presented for Pittsburgh No.8 bituminous coal. Outlet gas temperature was 515°C for BGC/L and 675°C for Lurgi dry-bottom gasifiers. Coal test of North Dakota lignite carried out in Lurgi gasifier showed about 30% reduction in steam consumption compared with gasification of Pittsburgh No.8 coal in this type of gasifier*, i.e., for processing of lignite around 2.5 kg of steam per kg of MAF coal may be assumed. Also, oxygen consumption would be lower.

Analysis of the data presented in Table 4-2 shows that the methane content in the raw gas leaving Lurgi dry-bottom or BGC/L fixed-bed gasifiers has the highest level (6.1 to 13.1 vol%). Raw gas generated in the fluidized-bed gasifiers has lower methane content (3.0 vol% for HTW and 4.3 vol% for Westinghouse gasifiers). The raw gas leaving the entrained-flow gasifiers,

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* Direct communication with process developer, 1982.

4.1 FACTORS AFFECTING SYNGAS PRODUCTION FROM COAL AND SUITABILITY OF SASKATCHEWAN LIGNITES FOR SELECTED PROCESSES (cont'd)

4.1.5.1 Methane Content (cont'd)

Texaco or GKT, contains negligible amounts of methane (0.1 vol%).From this point of view, entrained-flow gasifiers are most suitable for production of synthesis gas in an indirect coal liquefaction complex when maximum liquid products are desired.

4.1.5.2 Liquid Byproducts

Due to relatively low outlet-gas temperature in the case of fixed-bed gasifiers, the raw gas will contain condensible byproducts (tars and heavy oils, which amounts to roughly 5% of the coal feed), while the gas produced in other types of gasifiers (HTW, Westinghouse, Texaco and GKT) is essentially free from liquid byproducts.

The liquid byproducts produced in fixed-bed gasifiers can be sold or recycled. However, the equipment to separate the byproducts increases the capital and operating costs.

4.1.5.3 Hydrogen-to-Carbon-Monoxide Ratio

The ratio of hydrogen to carbon monoxide, H_2/CO, in the synthesis gas to be used in an indirect coal liquefaction complex (F-T synthesis, Mobil MTG process or for methanol synthesis) has to be around 2.0 to 2.5. If H_2/CO in the gas is lower, shift conversion of a portion of CO to H_2 will be required. Little "shifting", if any, is required for the gas produced from Lurgi fixed-bed gasifier, while raw gas leaving fluidized-bed or

Table 4-2 Typical Composition of Raw Gas and Operating Parameters

ITEM	Gasifiers (Oxygen Blown)					
	Fixed Bed		Fluidized Bed		Entrained Flow	
	Lurgi	BGC/L	HTW	W	Texaco**	GKT
Low Rank Coal Processing (lignite, brown coal)	Yes	No	Yes	No	Yes	No
Gas Composition, vol%						
CO_2	34.0	3.3*	N/A	12.8	34.8	11.9
CO	14.4	56.4*	N/A	53.5	35.0	55.9
H_2	36.2	29.1*	N/A	28.1	29.3	29.6
CH_4	13.1	6.7*	3.0	4.3	0.1	0.1
C_nH_m	0.9	0.7*	-	-	-	-
Others	1.4	3.8*	N/A	1.3	0.8	2.5
H_2/CO	2.5	0.5*	-	0.5	1.2	0.5
Pressure, MPa	2.8*	2.8*	1.0	3.0	9.1	0.05
Temp., °C (max)	1000*	1500*	990	945	1900	1900
Steam kg/kg MAF coal	3.5*	0.4*	0.9	0.2	-	0.1
Oxygen, kg/kg MAF coal	0.7*	0.6*	0.4	0.5	1.18	0.72

* Gas composition and operating parameters for bituminous coal Pittsburgh No.8.

** Data and parameters estimated by the process developer for gasification at Willow Bunch lignite on Texaco gasification process. All other data presented are for lignites (except for HTW gasifier processing Rhenish brown coal).

Source: G. Roberts et al, 14th World Gas Conference, May 1979.

4.1.5.3 Hydrogen-to-Carbon-Monoxide Ratio (cont'd)

entrained-flow considered will require shift conversion of a substantial portion of CO to H_2, thus increasing the capital and operating cost for raw gas treatment.

4.1.5.4 Carbon Dioxide Content

Table 4-2 shows that CO_2 content in the raw gas is much higher for gasification of lignites than for processing of bituminous coals (see column for BGC/L gasifier). This is due to usually higher oxygen content in lignites.

4.1.6 **Operating Pressure**

There is an advantage in gasifier operation at elevated pressure since capacity or output of the gasifier increases with pressure (as a rough rule-of-the-thumb, capacity increases as P). If the gasifier operating pressure is close to atmospheric, additional energy (and therefore increased operating costs) will be required for compression of gas since gas cleaning systems as well as the gas synthesis processes are usually carried out at elevated pressures. On the other hand, to manufacture and operate a pressure gasification unit require higher capital, operating and maintenance cost compared with the "atmospheric" gasification unit.

Among the gasifiers considered in the present study, only the GKT gasifier operates at close to atmospheric pressure.

4.1.7 **Oxygen and Steam Consumption**

The highest oxygen consumption is required for entrained-flow gasifiers (see Table 4-2). This is because of the high operating temperature that has to be maintained in this type of gasifier. The oxygen consumption in the Texaco gasifier is even higher (1.18 kg/kg MAF coal) than GKT (0.72 kg/kg MAF coal) due to the wet coal feeding in the case of the Texaco gasification process. Outside steam consumption (in addition to steam generated in the gasifier from water contained in the feed coal) is rather small (< 0.1 kg/kg MAF coal).

Fluidized-bed gasifiers, HTW and Westinghouse, have more moderate oxygen consumption (0.4 to 0.5 kg/kg MAF coal) compared with entrained-flow gasifiers, due to lower operating temperature. Outside steam consumption is beteween 0.2 to 0.9 kg/kg MAF coal (oxygen/steam ratio is 0.44 for HTW and 2.5 for Westinghouse gasifiers).

In the Lurgi dry-bottom gasifier, steam consumption has to be very high to keep the temperature on the rotating grate below the ash fusion point. By comparison, the steam comsumption in BGC/L slagging gasifiers is equal to about 12% of that required for operation of Lurgi dry-bottom gasifiers when processing the same coal, Pittsburgh No.8 (see Table 4-2). Therefore, the oxygen-to-steam ratio in this case is 0.2 for Lurgi and 1.5 for BGC/L gasifiers respectively. For lignite gasification in the Lurgi gasifier, steam consumption would be lower (around 2.5 kg/kg MAF coal as discussed earlier). However, absolute steam consumption would still be high in order to keep the temperature above the ash fusion point. Oxygen consumption would also be lower when processing lignites, as discussed earlier.

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4.2 RANKING OF COAL GASIFICATION PROCESSES FOR INDIRECT LIQUEFACTION OF DESIGNATED SASKATCHEWAN LIGNITES

Operating parameters and typical composition of raw gas produced in the gasifiers considered in this study was presented in Table 4-2. The status of these advanced coal gasification processes is summarized in Table 4-3.

Analyzing the properties of designated Saskatchewan lignites, and considering the status of each gasification process, its advantages/disadvantages, the following conclusions may be drawn:

The Lurgi gasifier (Figure 4-2)is a first generation pressurized, fixed-bed, dry-bottom gasifier which has been commercially proven on high-ash-content South African lignites (27 wt% ash on MF basis). Also, North Dakota lignite has been successfully tested in the Lurgi gasifier in a large-scale test carried out by SASOL in South Africa* (35 wt% moisture on as-received basis, 10.2 wt% ash on MF basis).

If the Lurgi gasifier is selected for syngas production, high amounts of methane in the raw gas will be produced and some liquid hydrocarbon byproducts may be expected. In order to maximize the syngas production, methane must be separated and reformed to syngas. An alternative is to use the methane separated from raw gas as a fuel. The additional equipment required to separate methane and to reform it to syngas (if required) would result in higher capital and operating costs. Also, additional equipment is required for separation of liquid byproducts from the raw gas.

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* Direct communication by J.A. Mikhlin with SASOL, South Africa, August 1982.

Table 4-3 Status of Advanced Coal Gasifiers for Syngas Production

Process	Status	Development Prospects
Lurgi	100 Commercial Units max 100,000 m^3/h gas (D 5m)	Commercially Proven
Koppers-Totzek	79 Commercial Units max 50,000 m^3/h gas (4 burners)	Commercially Proven
British Gas/Lurgi	350 t/d coal (D 1.85m); 500 t/d coal (D 2.45m), Westfield, Scotland	Eng. Design for Florida Power, U.S.A 1000 t/d coal, (D 3m) X 3 units
Texaco	- 15-20 t/d, Montebello, U.S.A. - 165 t/d coal, Oberhausen-Holten, F.R.G.	- TVA Coal-to-Methanol Plant, U.S.A - Cool Water Coal Gasification (1000 t/d coal) - Tennessee Eastman Project, U.S.A
HT Winkler	- 32 t/d coal, Freschen, Germany	- 1400 t/d MF coal (37,000 m^3/h gas) by 1983, F.R.G. - 1400 t/d x 4 units (147,000 m^3/h gas) by 1990, F.R.G.
Westinghouse	- 24 t/d coal, Waltz Mill, PA - Eng. Design of Commercial-Prototype Gasifier, 1200 t/d coal	- 1200 t/d coal at SASOL-II (by 1986), S. Africa.

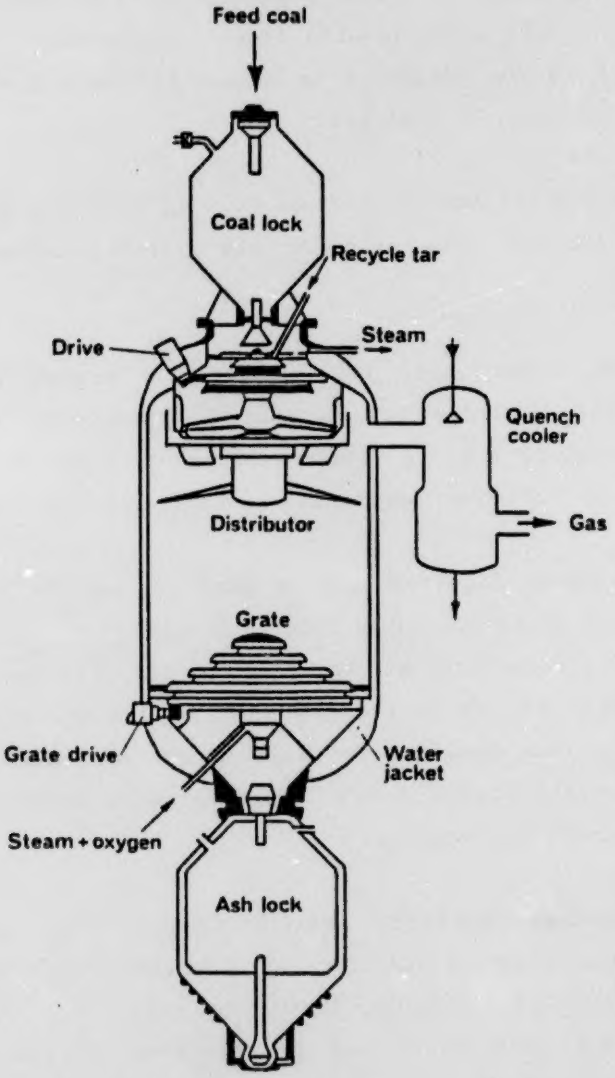

Figure 4-2 Lurgi Dry-Bottom Gasifier

4.2 RANKING OF COAL GASIFICATION PROCESSES FOR INDIRECT LIQUEFACTION OF DESIGNATED SASKATCHEWAN LIGNITES (cont'd)

Large amounts of fines expected in the Saskatchewan lignite (around 25%) would require fines separation. Coal fines may be burned in the utilities or briquetted to a desired size before feeding into the gasifier.

A substantial amount of steam (2.5 kg/kg MAF coal) is required to keep the ash temperature in the gasifier below the ash fusion point.

On the other hand, the commercially proven Lurgi dry-bottom gasifier is robust and stable in operation. Also, little CO shifting, if any, is required before feeding the cleaned raw gas into the subsequent synthesis reactor.

Saskatchewan lignites may be gasified in the Lurgi dry-bottom gasifier after the extra fines are separated. However, to define actual raw gas composition, coal devolatilization under pressure, behaviour of ash in the gasifier, oxygen and steam consumption, etc., a coal gasification test using representative samples of Estevan and Willow Bunch lignites, as requested by the process developer*, is required.

The British Gas/Lurgi gasifier (Figure 4-3) is a fixed-bed, pressure, slagging gasifier. The British Gas/Lurgi gasifier is a derivation of the Lurgi dry-bottom gasifier. The main purpose for this development was gasification of the unreacted high volatile bituminous coals, particularly those with low melting point of the ash and which showed poorer performance in the dry-bottom Lurgi gasifier. BGC/L was never tested on lignites.

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* Direct communications with process developer, March 1981 (see Appendix C-6; and Shaunavon Coal Utilization Study, Phase II, Volume II, SPC report prepared by SASKMONT Eng., July 1980.

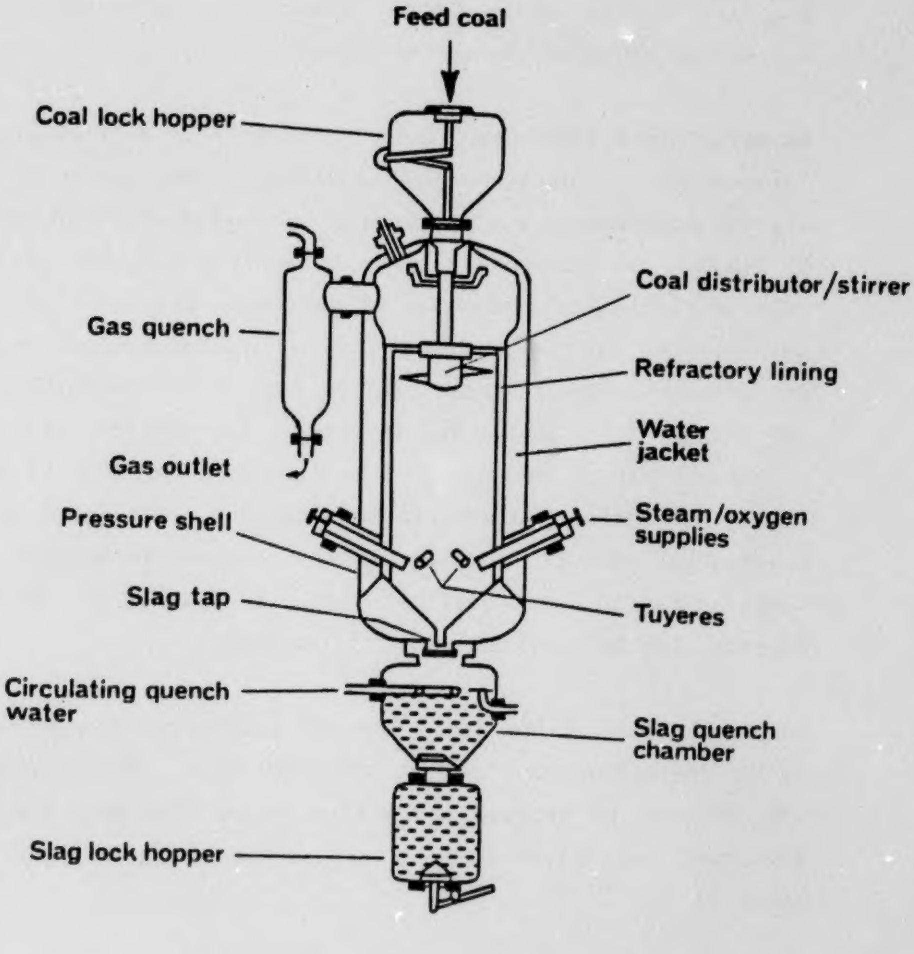

Figure 4-3 British Gas/Lurgi Slagging Gasifier

The process developer, being contacted by SNC, requested a pilot-plant test at Westfield to assess the suitability of Estevan and Willow Bunch lignites for processing in British Gas/Lurgi gasifier (see Appendix C-7). However, a pilot-plant test is beyond the scope of the current study.

Koppers-Totzek (GKT) gasifier (Figure 4-4) is a first generation "atmospheric", entrained-flow gasifier commercially proven on lignite predried to 8 wt% moisture and having up to 40 wt% ash on MF basis. As discussed earlier in Section 4.1, the gasifier is quite suitable for production of synthesis gas since practically methane-free gas is produced. Also, no condensible byproducts are present in the raw gas. At the same time, the H_2/CO ratio in the raw gas is around 0.5. Therefore, substantial portion of CO in the gas has to be converted to H_2 before feeding the gas into the F-T or methanol synthesis for indirect coal liquefaction. A substantial amount of oxygen is also required to maintain a high temperature in the gasifier (around 0.7 kg/kg MAF coal). However, little external steam is required.

Pulverized coal is fed into the GKT gasifier. Therefore, there is no limitation on fines in the coal feed. Based on coal and ash analyses of Estevan and Willow Bunch lignites, the process developer estimated these lignites as suitable (see Appendix C-8).

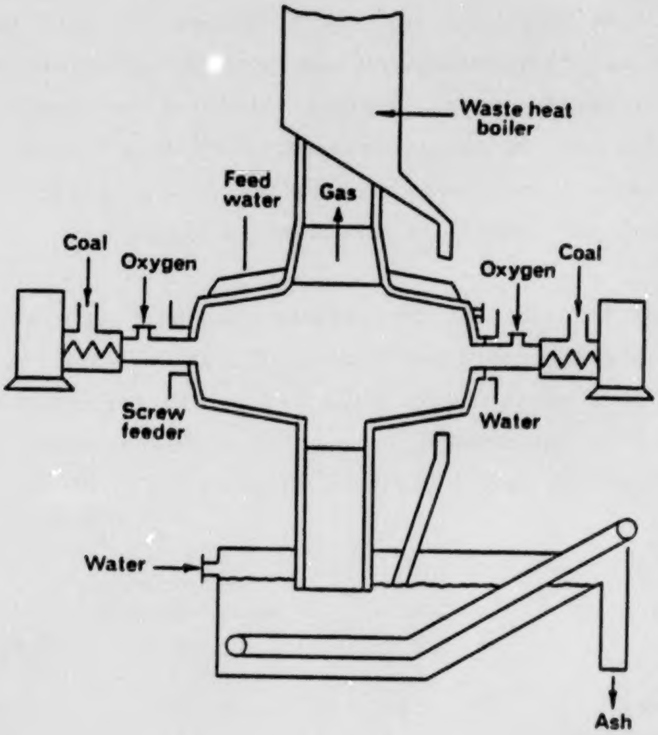

Figure 4-4 Koppers-Totzek (GKT) Gasifier

4.2 RANKING OF COAL GASIFICATION PROCESSES FOR INDIRECT LIQUEFACTION OF DESIGNATED SASKATCHEWAN LIGNITES (cont'd)

The main concern in application of an "atmospheric" GKT process in comparison to pressure, entrained-flow gasifiers (e.g., Texaco and Shell) is the process economics (see discussion on Operating Pressure, Section 4.0). However, the recent Krupp-Koppers study* indicates that the overall economics of coal gasification in entrained-flow atmospheric and pressure gasifiers depends on cost and properties (e.g., heating value) of the feed coal. The plot of the cost of syngas produced (DM/1000 m^3 of CO + H_2) in both atmospheric and pressure entrained-flow gasifiers versus the cost of coal per (DM/GJ) is presented in Figure 4-5.

It may be seen that the product cost of syngas is favourable for atmospheric, entrained-flow gasification when the coal cost is less than or equal to DM/GJ 1.5 to 2.5 (or about Can. $/GJ 0.75 to 1.28). At present, the price of Saskatchewan lignites is in the range of Can. $ 0.6 to 0.75/GJ as shown below:

	Can $/ tonne**	GCV, GJ/tonne as-received coal	Can $/GJ
Estevan	9.80	15.78	0.62
Willow Bunch	9.80	13.25	0.74

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* H. Staege, Entrained-Bed Gasification, "Hydrocarbon Processing", March 1982.

** Letter from Dr. F. Mourits, Coal Liquefaction Study Management, Saskoil, to Dr. J.A. Mikhlin, SNC Inc., January 1982.

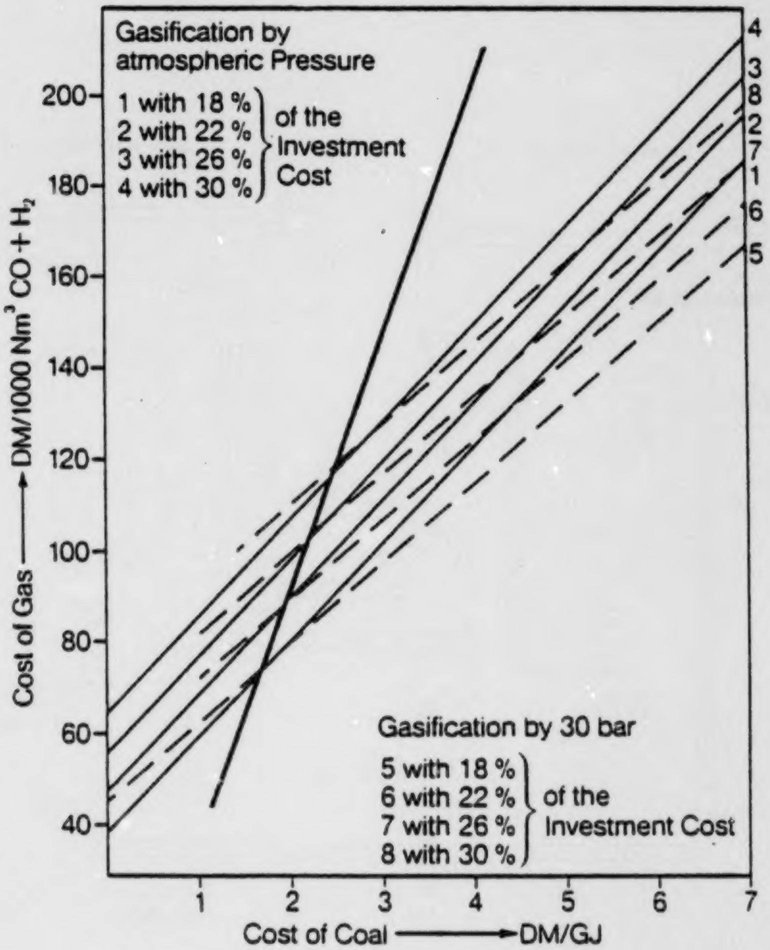

Figure 4-5. Product Cost of Gas Generated in Koppers-Totzek Gasifier versus Cost of Coal

Source: H. Staege, Entrained-Bed Gasification "Hydrocarbon Processing" March 1982

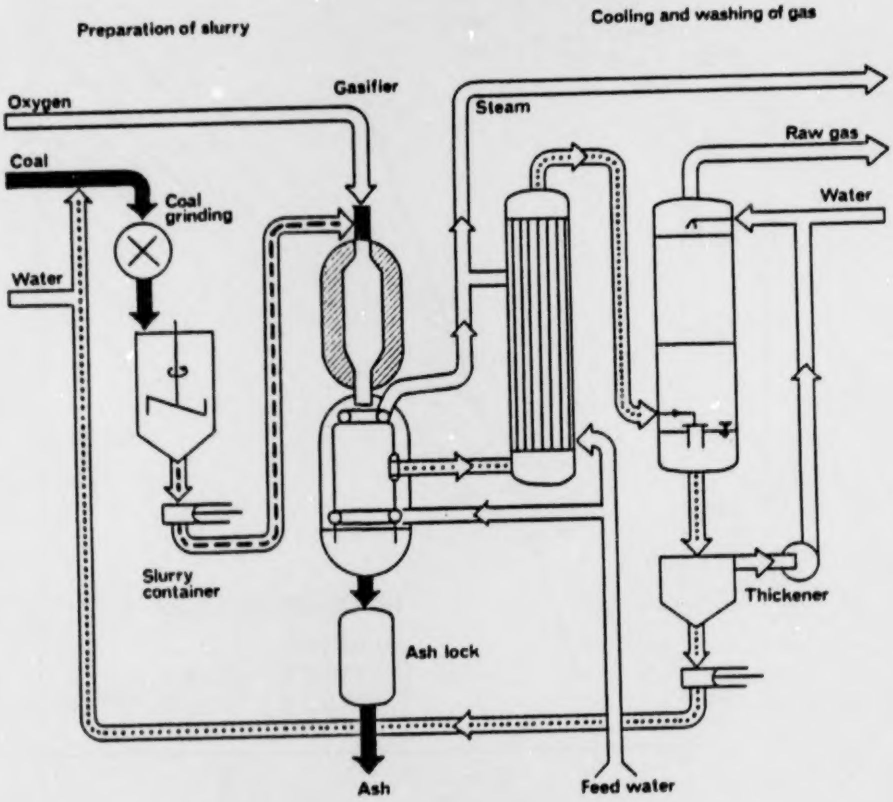

Figure 4-6. Texaco Coal Gasification Process (Waste Heat Recovery Mode)

4.2 RANKING OF COAL GASIFICATION PROCESSES FOR INDIRECT LIQUEFACTION OF DESIGNATED SASKATCHEWAN LIGNITES (cont'd)

So, it may be assumed that the Koppers-Totzek coal gasification process using Estevan or Willow Bunch lignites as a feedstock is not only technically proven on a commercial scale but would be also economically viable.

The Texaco coal gasification process (Figure 4-6) is a second generation pressurized, entrained-downflow, slagging gasifier. The Texaco coal gasifier is refractory lined but is similar to the commercially proven Texaco oil gasification process for generating synthesis gas from materials ranging from residual oils to naphtha.

As in the GKT process, due to high temperature maintained in the Texaco gasifier, methane and condensible byproducts are negligible. In addition, the pressurized Texaco gasifier would have higher specific gas output compared with the "atmospheric" entrained-flow GKT gasifier. However, suitability of Saskatchewan lignites for gasification in the Texaco gasification process is doubtful because of restrictions in preparation of concentrated coal slurry in water to be fed in the gasifier (see 4.1.2).

HTW gasifier (Figure 4-7) is a fluidized-bed pressurized gasifier, a derivation of the first generation "atmospheric" Winkler gasifier. While the "atmospheric" Winkler gasifier is commercially proven on high moisture, high ash lignite similar to Saskatchewan lignites the HTW is being extensively demonstrated on low ash (3 wt% MF basis) Rhenish brown coal only.

The complete raw gas composition has not been reported by the process developer. However, the methane content of about 3.0 vol% in the raw gas reported by the process developer (see Table 4-2) may be seen as tolerable for syngas production. Also, absence of condensible byproducts in the raw gas is the positive factor. Oxygen consumption is much lower in comparison to an entrained-flow gasifier (0.4 kg/kg MAF instead of 0.72 to 1.18 kg/kg MAF coal, see Table 4-2).

The specific gas output of HTW gasifier (m^3/m^2h) was reported to be 2.3 times higher in comparison to the "atmospheric" Winkler gasifier and the carbon conversion is 95% instead of 91%. On the other hand, steam consumption in the HTW gasifier is 2.4 times higher than in the atmospheric Winkler gasifier, while oxygen consumption remained at approximately the same level.

Ground predried coal is fed to the HTW gasifier. Therefore, there is no limitation on fines in the feed coal. As stated by the process developer, the coal to be used in the HTW coal gasification process must have sufficiently high reactivity to steam below the ash fusion point. The ash fusion temperature has to be equal to or higher than 1000°C.

According to these requirements, Saskatchewan lignites are, in principle, suitable feedstocks for the HTW process. However, the actual experimental data on gasification of coals similar to Estevan or Willow Bunch lignites in the HTW process are not available.

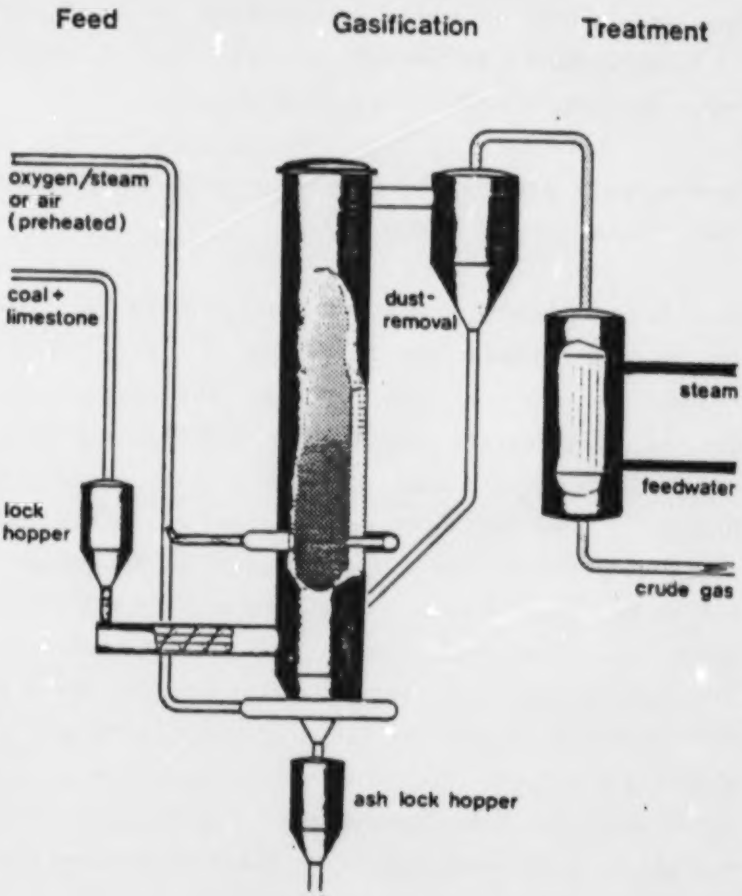

Figure 4-7. High-Temperature Winkler Process

The general suitability of Saskatchewan lignites for processing in the HTW process was confirmed by Rheinbraun based on coal and ash analyses (see Appendix C-3). At the same time, the process developer requested to carry out a coal test in a 50 kg/h bench-scale unit at Aachen to determine the gasification behavior of these lignites in the HTW process. Such a test is beyond the scope and budget of the current study.

Westinghouse gasifier (Figure 4-8) is a single-stage, pressurized, fluidized-bed gasifier.

Both HTW and Westinghouse are fluidized-bed gasifiers, operating in the same temperature level (945 to 990°C) and the methane content in the raw gas is also similar (3.0 to 4.3 vol%). However, the reactor design, the coal feeding system, and ash disposal system are quite different. The operating pressure is higher in the Westinghouse gasifier but steam consumption is lower compared to the HTW gasifier (3.0 MPa versus 1.0 MPa and 0.2 kg/kg MAF coal versus of 0.9 kg/kg MAF coal respectively, see Table 4-2). Several lignites have been tested in the 24-t/d Westinghouse coal gasification pilot plant. Raw gas composition and operating parameters for a lignite (9.96 wt% ash MF basis) predried to 15 wt% total moisture is presented in Table 4-2. The H_2/CO ratio of 0.5 suggests that a sufficient portion of CO in the gas must be converted to H_2 prior to feeding the syngas into the F-T or methanol synthesis reactor.

At present, Westinghouse Electric Corporation is developing a 1200-t/d gasification demonstration unit to be installed at SASOL-II complex in South Africa. The unit is planned to be on stream by mid-1986.

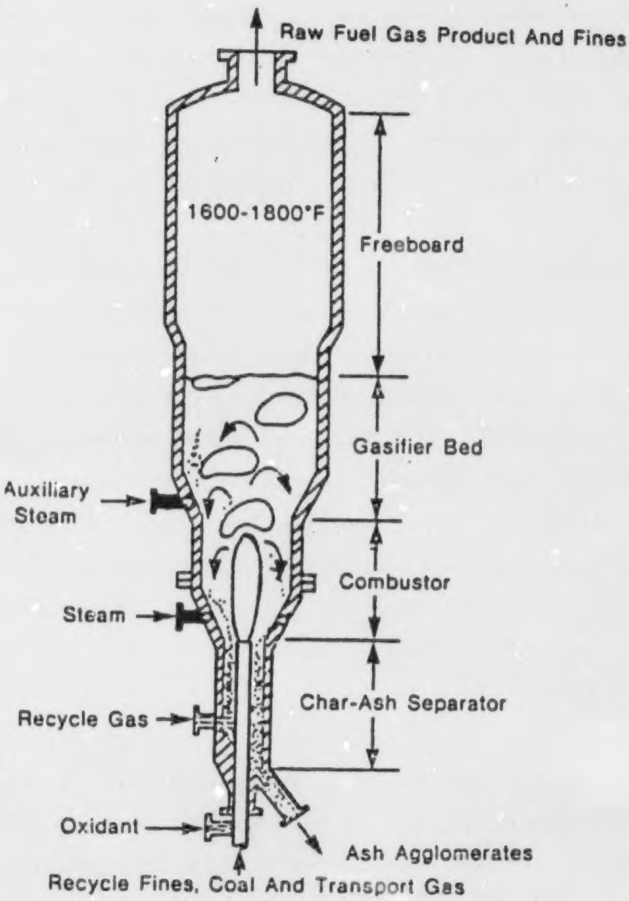

Figure 4-8. Westinghouse Gasifier

4.2 RANKING OF COAL GASIFICATION PROCESSES FOR INDIRECT LIQUEFACTION OF DESIGNATED SASKATCHEWAN LIGNITES (cont'd)

Based on the process developer's experience in testing of lignites in the Westinghouse coal gasification process, it may be considered that the Saskatchewan lignites are suitable feedstocks for the Westinghouse gasifier.

The process developer confirmed the general suitability of Saskatchewan lignites in their process. However, to determine the quantitative process data, Westinghouse Electric Corporation requested to carry out a preliminary coal test (see Appendix C-9) which is beyond the scope and budget of the current study.

In summary, it may be concluded that the GKT coal gasification process including the atmospheric entrained-flow Koppers-Totzek gasifier appears to be most suitable for gasification of designated Saskatchewan lignites, as a single gasification process to be selected for further techno-economic assessments of the indirect coal liquefaction alternatives considered (Stage II of the Study). This process is commercially proven on high-ash lignites, has no restrictions on processing of fines and contains no methane in the product gas. Also, application of this process may be economically viable for the low-cost Saskatchewan lignites.

The second-generation fluidized-bed gasifiers, HTW and Westinghouse, appear to be the next best suitable. For a future program on indirect liquefaction of Saskatchewan lignites, a bench-scale coal gasification test, and subsequent techno-economic evaluation of these pressurized fluidized-bed gasifiers is recommended for a comparative evaluation with the entrained-flow Koppers-Totzek gasifiers.

4.2 RANKING OF COAL GASIFICATION PROCESSES FOR INDIRECT LIQUEFACTION OF DESIGNATED SASKATCHEWAN LIGNITES (cont'd)

The coal gasification tests are necessary for adjusting the predicted results (operating conditions, raw gas composition and specific gas output) by the process developers as well for defining the "gasification behavior" of coal ash and other factors affecting the overall process performance and economics.

If lignite conversion to SNG is considered, more detailed techno-economic evaluation of Lurgi gasification process is recommended.

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COAL PYROLYSIS

5.0 OVERVIEW OF ADVANCED PYROLYSIS PROCESS AND ITS SUITABILITY FOR SASKATCHEWAN LIGNITES

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5.0 COAL PYROLYSIS: OVERVIEW OF ADVANCED PYROLYSIS PROCESS AND ITS SUITABILITY FOR DESIGNATED SASKATCHEWAN LIGNITES

The Lurgi-Ruhrgas (L-R) process is the only advanced pyrolysis process selected by the Study Committee for evaluation of the designated Saskatchewan lignites. A comprehensive technical assessment including process history and development, relevant experimental results pertaining to processing of lignites , overall configuration of a conceptual commercial plant, problem areas and commercialization constraints, etc., are described in Volume II of the report.

In this summary volume (Volume I), a brief overview highlighting only the specific features of the process, its status, and its suitability for the designated Saskatchewan coals (Estevan and Willow Bunch lignite) is presented.

5.1 LURGI-RUHRGAS PROCESS

The Lurgi-Ruhrgas (L-R) process is a flash carbonization process developed by Lurgi Kohle und fur Mineralöltechnik GmbH, in cooperation with Ruhrgas AG. The process has been used for carbonization of coals, oil shales, tar sands and asphaltic rocks. Cracking of hydrocarbons to produce olefins has also been tested.

The novelty of the L-R process is the use of hot, fine-grained product char as a heat carrier which when intimately mixed with the feed coal in a mechanical screw-mixer/carbonizer results in carbonization of the coal. Another essential feature is that char is simultaneously heated and pneumatically conveyed to the carbonizer thus eliminating the need for a separate heater.

5.1.1 **Process Description**

A simplified process flow diagram of the L-R process is shown in Figure 5-1. Dried and pulverised coal (2 wt% moisture, 5 mm size) from the coal preparation section is fed into the screw-mixer/carbonizer where it is intimately mixed with hot char at about 700°C from the collecting bin. The coal-to-char mixing ratio is kept between 4 and 8 to maintain the mixing temperature of approximately 600°C.

The residence time in the mixer/carbonizer is kept very short to suppress secondary decomposition reactions and thus allowing a higher liquid product yield. The gases and condensible vapors leaving the mixer/carbonizer are sent to a cyclone where the entrained coal and char particles are removed and returned to the circulation system downstream of the surge bin. The gases/vapors are sent to the gas/liquid separation section.

The freshly produced char and the cooled heat-carrier (recycled char) are discharged into a surge bin from where the solids are withdrawn and pneumatically conveyed by hot combustion gases to the carbonizer/mixer. The flue gases are sent to a waste-heat recovery system (not shown in the diagram).

Tar obtained from the gas/liquid separation is a non-distillable, highly viscous hydrocarbon liquid which is hydrogenated in a catalytic reactor to obtain distillate products. Hydrogen, as shown in the diagram, is provided by steam reforming of hydrocarbon gases obtained from the gas plant. Alternatively, hydrogen may be produced via gasification of char. For Saskatchewan lignites, hydrogen is assumed to be produced by the latter method.

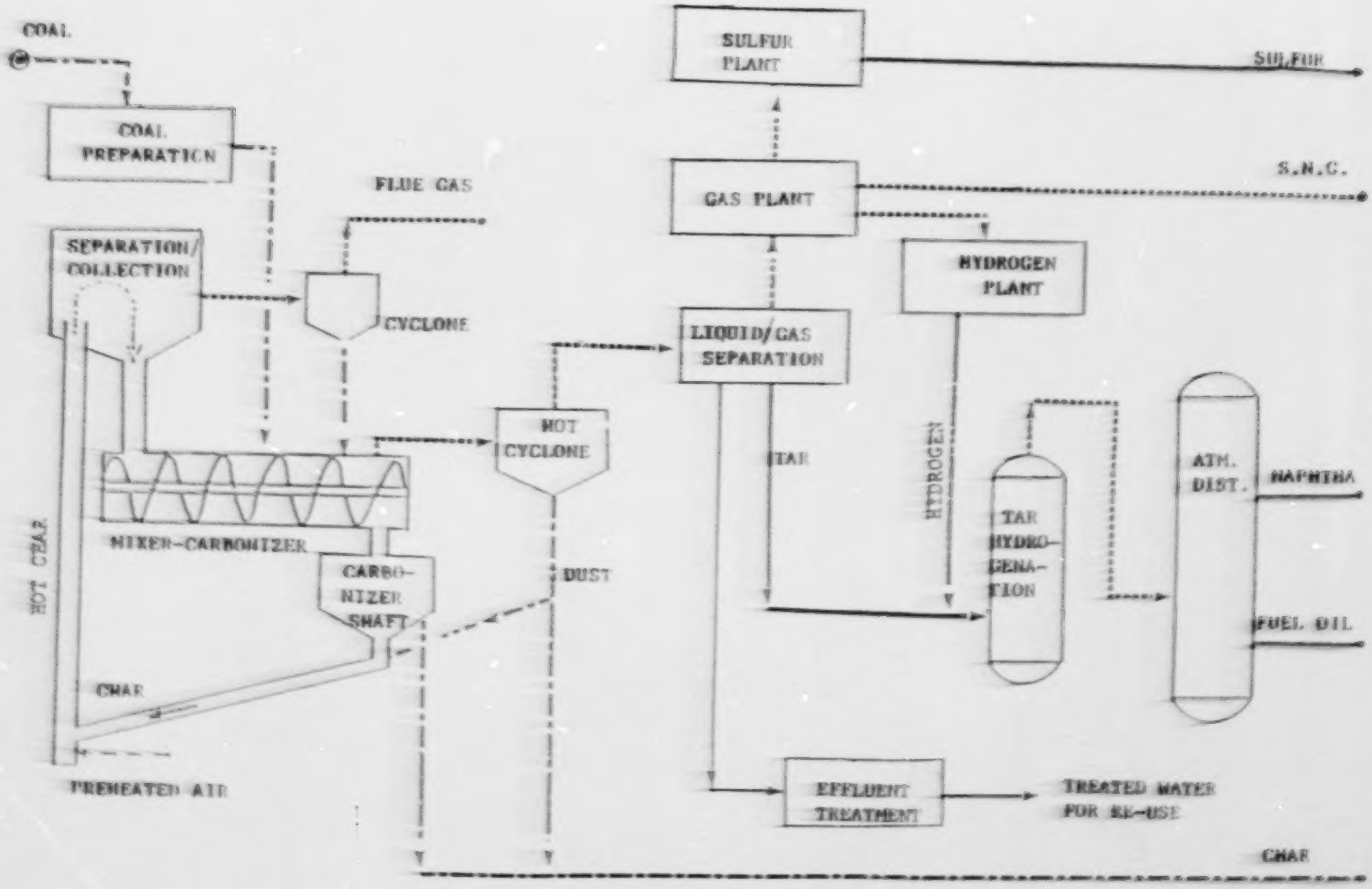

Figure 5-1 Overall Conceptual Configuration of the L-R Process

5.1.2 Product Yield and Efficiencies

The L-R process has been tested on various types of coals ranging from lignites to high volatile bituminous coals. The largest commercial L-R process plant at is in operation in Yugoslavia processing 1600 t/d of lignite. (36.0 wt% moisture, 8.6 wt% ash MF coal). A German subbituminous coal has also been extensively tested in the pilot plant. Analysis of this coal is given in Appendix B.

A Fischer Assay of 10 wt% tar on an MAF basis and liquid product yields of 190% of Fischer Assay are assumed. The product distribution for Estevan lignite was predicted using German subbitumious coal as a reference coal. Make-up hydrogen and additional utility requirements for self-sufficient operation is met by utilization of char. The amount of char shown in Table 5-1 is about 70% of the total char produced. The liquid product yields and efficiencies have been adjusted to zero electricity input using the thermal factor of 10,000 Btu/kWh (10,550 kJ/kWh) and a fuel oil equivalent factor of 6.00 MM Btu/bbl (6.33 GJ/bbl). The adjusted liquid product yields are 1.0 bbl/t MF coal or 1.2 bbl/t MAF coal. Similarly, the overall and liquid product efficiencies are 72.4% and 31.4%, respectively.

In terms of process inputs, the L-R commercial plant operating with Estevan lignite will require about 54 MW electricity and 190 L/s as make-up water.

Table 5-1 Estimated Product Yield and Efficiencies of a Conceptual L-R Plant

[Basis: 25,000 t/d as-received coal]

	Estevan Lignite
Coal (as received)	
Moisture, wt%	35.0
Ash, wt% of MF coal	15.4
Coal Fed to Reactor	
Moisture, wt%	2.0
Ash, wt% of MF coal	15.4
Input	
MF Coal to Reactor, t/d	16250
MF Coal to Gasifiers, t/d	-
MF Coal to Offsites, t/d	-
Elctricity, MW	54
Output, t/d (bbl/t)	
Fuel Gas	1710
Propane/Butane	-
Naphtha (C_5-200°C)	1526 (11187)
Fuel Oil (200 - 525°C)	1107 (7579)
Char*	5050
Sulfur	34
Liquid Product Yield** (C_5-525°C)	
bbl/t as-received coal to plant	0.66
bbl/t MF coal to plant	1.02
bbl/t MAF coal to plant	1.20
Naphtha in Total Liquid, wt%	58.0
Thermal Efficiency, %	
Overall	72.4
Liquid Products	31.4

* About 30% of the total char produced is used for make-up hydrogen and utilities production. The value shown here is the net char yield, i.e., after utilization for hydrogen production.

** Product yields and efficiencies have been adjusted to zero electricity import using a thermal conversion factor of 10,000 Btu/kWh (10,550 kJ/kWh) and a fuel oil equivalent factory of 6.0 MM Btu/bbl (6.33 GJ/bbl).

Source: R.W. Rammler "The Lurgi-Ruhrgas Process for Carbonization of Coal" paper presented at the EPRI Conference on coal Pyrolysis, Palo Alto, California, Feb. 25-26, 1981.

5.1.3 Process Status

Two commercial L-R units with a throughput capacity of 1600 t/d of dry lignite have been built and commissioned in Yugoslavia since 1963. The plants were built to produce fine-grained char of 15 to 20 wt% volatile-matter content for use as an additive to coking coal in a conventional coke oven works. These plants have been shut down. Two other L-R units with feed coal capacities of 350 and 700 t/d were built at Bottrop (Germany) and Scunthorpe (Great Britain) for production of char fines from bituminous coals for hot briquetting. These units were operated at carbonizing temperatures around 750°C.

Renewed interest in the L-R process for partial liquefaction of coal has led Lurgi to build a new pilot plant test unit of a feed capacity up to 24 t/d on their own premises in Frankfurt/Main, F.R.G. The purpose of the unit is to test feedstocks of various types and origins in addition to coal.

The carbonization section of the L-R plant can be regarded as a proven commercial operation. Similarly, the hydrogen production plant using hydrocarbon gas or char as feedstock is also commercially proven for many years in the chemical industry notably in production of synthetic ammonia. However, the tar hydrogenation to yield synthetic crude has not been demonstrated commercially.

Based on above information, it may be expected that the L-R process may be commercialized by 1990.

5.1.4 Suitability for Saskatchewan lignites

The suitability of the Lurgi-Ruhrgas process for different types of lignite has been demonstrated at various levels of operation from 240 t/d pilot plant to 1600 t/d commercial plant. The ash and moisture content in subbituminous and lignitic coals varied from 5 to 28 wt% and 2 to 36 wt% respectively indicating the suitability of Saskatchewan lignites* in the L-R process.

Nevertheless, the process suitability must also be viewed on the basis of desired end-products. If char is considered as a main product, then the process appears to be suitable for most types of coal including Saskatchewan lignites. On the other hand, if the syncrude (naphtha, fuel oil) are desired as the main products, then the preliminary judgement on process suitability may be made on the basis of Fischer Assay tar yields of the feed coal. Coals with Fischer Assay tar yield less than 10%, as indicated by the process developer, should not be considered as suitable for production of liquid products.

The Fisher Assay of Estevan and Willow Bunch lignite, as determined by B.C. Research Laboratories, are presented in Table 5-2. The Estevan lignite has almost 10 wt% Fischer Assay tar yield but the Willow Bunch lignite showed only 6 wt% tar yield. Char yield in both cases is in the range of 65 to 70 wt% MAF coal.

Since the present study is directed towards evaluation of processes for production of mainly transportation fuels, the L-R process is considered as not suitable for Willow Bunch lignite. For Estevan lignite, the L-R process appears to be marginally suitable only

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* Moisture 35.0 wt% in as-received coal, ash 15-21 wt% MF coal

5.1.4 **Suitability for Saskatchewan lignites** (cont'd)

if the char is effectively utilised. Char is not a desired product for this study (see product criteria, Chapter 2-2), but combustion of char in large power plants may be viewed as one of the ways to utilize the surplus char product.

Since Willow Bunch lignite is the only coal feedstock designated by the Study Committee for techno-economic assessment, the L-R process is not considered for further evaluation (Stage II of the Study).

Table 5-2 Fischer Assay of Saskatchewan Lignites as Determined by B.C. Research Laboratories

[Carbonization Temperature = 500°C]

	Estevan Lignite		Willow Bunch Lignite
	Untreated	Predried to 10 wt% H_2O	Untreated
Fischer Assay Yield, wt% MAF coal			
Char	65.0	69.4	65.8
Tar	9.4	9.4	5.8
Water Formed	10.1	9.6	15.5
Total Gas (by difference)	15.5	11.6	12.9
Composition of Gas, vol%			
CH_4	17.5		13.8
C_2H_6	3.6		2.2
H_2	28.0		44.9
CO	9.0		4.2
CO_2	41.9		34.9
H_2S	0.1		0.1

Source: Liquefaction and Pyrolysis Tests at Samples of Estevan and Coronach Lignites prepared for SNC Inc., B.C. Research Laboratories, Project Report 2-02-691, 1982.

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6.0 TECHNO-ECONOMIC ASSESSMENT OF PRESELECTED PROCESSES

6.0 TECHNO-ECONOMIC ASSESSMENT OF PRESELECTED PROCESSES

An overview and prescreening of direct and indirect coal liquefaction and pyrolysis processes were presented earlier in Chapters 3, 4 and 5 respectively. The following processing alternatives were selected for further techno-economic assessment.

- Direct Coal Liquefaction:
  - Liquid Solvent Extraction (LSE) process

- Indirect Coal Liquefaction:
  - SASOL-type Fischer-Tropsch synthesis with Synthol reactor.
  - Mobil Methanol-to-gasoline (MTG) process.
  - Production of pure methanol for transportation fuel or for blending with gasoline.

For indirect coal liquefaction alternatives, a Koppers-Totzek entrained-flow gasifier configuration for syngas generation was selected.

As mentionned earlier, the pyrolysis route was discarded from further techno-economic assessment, because the only process considered in the evaluation (L-R process) was found to be unsuitable for Willow Bunch lignite.

The technical assessment for each of the four "preselected" process alternatives is based on a conceptual design of a self-sufficient plant processing 25,000 t/d of Willow Bunch lignite and includes overall material balance and utilities requirements. The economic assessment include capital, operating and production cost estimates. The details of techno-economic assessments are presented in Volumes II and III for direct and indirect liquefaction processes respectively.

6.0 TECHNO-ECONOMIC ASSESSMENT OF PRESELECTED PROCESSES (cont'd)

In this volume, a brief description of the capital, operating and production cost estimation procedure and the results are given in the following pages. A summary of techno-economic results for final comparative evaluation and selection of the most promising process(es) is given in Chapter 6.2.

6.1 CAPITAL, OPERATING AND PRODUCTION COST

Preliminary capital, operating and production costs of each of the preselected processes have been made considering a self-sufficient, integrated plant processing 25,000 t/d of Willow Bunch lignite. These costs have been developed in accordance with the criteria described earlier in Chapter 2.6 and following the procedure approved by the Study Committee.

6.1.1 Capital Cost Estimate and Total Investment

The capital cost estimate and investment data available from the process developers/licensers and open literature for a different plant locations have been adjusted to the designated Saskatchewan area and Canadian dollars, as follows:

- Equipment items are assumed to be 25% Canadian content and 75% U.S. content. Items purchased in U.S. are not available in Canada, otherwise, a duty of 12.8% will be levied (Customs Canada, 27 Jan. 1982).

- All bulk materials (pipings, fittings, etc.,) are considered as 100% Canadian content, comprising 45% labour and 55% material components.

- The U.S. material and equipment content (75% of the total) is converted to Saskatchewan base by multiplying by a factor of 1.19 composed strictly of currency exchange rate (Royal Bank 27-01-82).

- The Canadian material and equipment content (25% of the total) is obtained by multipying the cost given for a

6.1.1 **Capital Cost Estimate and Total Investment** (cont'd)

U.S. location by a "material plus currency differential factor" of 1.379 (see Appendix D-3).

- Construction manpower/labor costs for a reference location is converted to Saskatchewan location by multiplying by a productivity adjustment factor of 1.42 (see Appendix D-4).
- Capital cost estimates is adjusted for inflation and converted to the base year (mid-1982) using the Chemical and Mineral Plant Price Index given in Appendix D-5.

Following the above procedure, the capital cost estimate for each of the preselected processes by individual processing area/units is calculated. The detailed breakdown for the LSE process is given in Volume II and for indirect processes in Volume III of the study report. However, the installed-plant cost with other investment components are shown in Table 6-1.

The installed-plant cost includes all items such as equipments, bulk material, labor, process contingency, licenses and prepaid royalties, engineering and support and other construction costs. A project and estimation contingency of 10% is added to cover uncertainties in cost estimation, equipment damage during construction, delay in equipment delivery and project schedule.

Cost of initial catalysts, chemicals and operating supplies comprises the cost of "initial charge" of these materials to the plant. These costs have been assumed to be equivalent to 1-month supply of catalysts, chemicals and consumables (except water) based on full plant-capacity.

Table 6-1 Breakdown of Total Capital Investment for Pre-Selected Coal Liquefaction Processes, in Millions of Canadian Dollars (mid-1982)
[Capacity: 25,000 t/d of "as-received" Willow Bunch Lignite]

Processes	LSE	Mobil MTG	Sasol Type F-T Plant	Methanol Plant
Installed-Plant Cost (including process contingencies)	2,028.37	1,743.4	2,005.67	1,517.10
Project and Estimation Contingencies	202.84	174.3	200.57	151.71
Total Plant Installation Cost (TPIC)	2,231.21	1,917.7	2,206.24	1,668.81
Licences and Prepaid Royalties	*	*	*	*
Initial Catalysts, Chemicals and Operating Supplies Cost	1.07	1.5	1.55	1.40
Start-up and Other Costs	66.94	57.5	66.19	50.06
Working Capital	66.94	57.5	66.19	50.06
Land	0.09	0.08	0.10	0.08
Interest During Construction	272.75	178.48	205.35	155.33
Total Capital Investment, "C"	2,639.00	2,212.76	2,545.62	1,925.74

* Included in the installed-plant cost

Sources:

W.P. Syrett, NCB report to SNC Inc. DR-496/2, 7/3, January 23, 1982
H. Staege, Krupp-Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982
B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979.
M. Schreiner, Mobil Research and Development Corporation, DOE Report No. FE-2447-13, August 1978.
R. Salmon et al., Oak Ridge National Laboratory, Report No. ORNL/TM-7581, April 1981.

6.1.1 **Capital Cost Estimate and Total Investment** (cont'd)

Start-up and other costs are taken as 3% of total plant installed cost (TPIC) to cover for operating expenses during commissioning and start-up period. It also includes additional capital for modifications required to bring the plant to full capacity, associated processes and environmental studies, contractors' fees, owners engineering offices and cost for obtaining construction permits.

Working capital is assumed as 3% of TPIC. Land cost is calculated on a basis of \$990/hectares (\$400/acre*).

Interest during construction is based on capital spending schedules given in Appendix D-2. Investment tax credits for interest during construction is claimed in the year of occurrence.

The capital investment, however, does not include:

- The cost of coal mine, mine development and transportation facilities from the mine to the plant.
- Railways, roads and other transportation facilities outside the plant site.
- Product distribution pipelines.
- Water supply and pipelines outside the plant site.
- Housing, community facilities and associated infrastructure.

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* 1 acre = 4047 m^2.

6.1 CAPITAL, OPERATING AND PRODUCTION COST (cont'd)

6.1.2 Operating, Maintenance and Production Costs

Annual operating, maintenance and production costs for each of the four "preselected" processes assuming a plant service factor of 85% (310 operating days per year) at 100% design capacity (25,000 t/d of coal) is given in Table 6-2. For the LSE process, the costs are given for two cases (65% and 82% extraction reactor efficiency on an MAF basis).

Operating and supervisory labour costs are calculated using a typical manpower distribution and average salary scale given in Appendix D-6. The total manpower estimated by the NCB for the LSE plant is 900, consequently, the operating labour force is 306 (34% of 900). The total manpower estimated by Krupp-Koppers for a methanol plant is 292. For plant with a Mobil-MTG processing unit, an additional 10 operators are required*, totalling to 302. For plant based on Fischer-Tropsch synthesis, an additional 108 operators are required*, totalling to 410.

Operating supplies are estimated at 30% of operating labour cost. A labour burden totalling 35% of direct labour cost is added to cover fringe benefits, sick leave, social security, unemployment insurance and shift differentials, etc.

Maintenance is estimated at 3% of total plant installation cost (TPIC). This includes replacement materials and equipment items, maintenance labour, supervision, labour burden and other regular maintenance contracts.

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* M. Schreiner, Mobil Research and Development Corporation, DOE Report FE-2447-13, August 1978.

Tabl

Fixed expenses

Operating Labou
Operating Suppl
Supervisory Lab
Labour Burden
Maintenance
Property Taxes
Plant Overhead

Variable Expenses

Coal Feed
Catalysts and C
Make up Water
Waste Disposal
Running Royalti

Total Annual Opera

Less By Product C

Total Annual Net-

Capital Costs **

TOTAL

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* Does not inclu
Similarly, pla

** Capital cost f
for Mobil MTG,

Sources:

W.P. Syrett, NCB
H. Staege, Krupp-
B.C. McBeath, ESC
M. Schreiner, Mob
R. Salmon et al.

6-2 Annual Operating, Maintenance and Production Costs, in Millions of
Canadian Dollars (mid-1982)
[Capacity: 25,000 t/d of "as-received " Willow Bunch Lignite]
[310 operating days per year]

	LSE PROCESS				
	65% extraction MAF basis	82% extraction MAF basis	Mobil MTG	SASOL Type F-T Plant	Methanol Plant
*	6.88	6.88	6.80	9.22	6.57
es	2.06	2.06	2.04	2.77	1.97
ur*	1.89	1.89	1.86	2.46	1.80
	3.06	3.06	3.03	4.09	2.93
	66.94	66.94	57.53	66.19	50.06
nd Insurance	61.36	61.36	52.74	60.67	45.89
	32.64	32.64	27.29	32.26	24.16
	75.62	75.62	75.62	75.62	75.62
emical	10.80	10.80	15.70	16.65	14.50
	4.30	4.10	3.77	3.69	3.74
	2.00	2.00	2.12	2.12	2.12
s	-	-	-	-	-
ing Costs	267.55	267.35	248.50	275.74	229.36
dits	12.95	13.04	2.42	2.42	2.62
erating Costs	254.60	254.31	246.08	273.32	226.94
	294.40	294.40	253.56	292.68	220.64
	549.00	548.71	499.58	566.00	447.58

maintenance labour and supervision as these items are covered in maintenance cost.
management and administrative staff are included in plant overhead.

tor "k" as defined in Equation 2 (Chapter 6.1.3) is 0.11155 for LSE Process, 0.11456
.11497 for F-T Plant, and 0.11457 for methanol plant.

eport to SNC Inc. DR-496/2, 7/3, January 23, 1982
ppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982
, DOE Report No. FE-2468-44, April 1979.
l Research and Development Corporation, DOE Report No. FE-2447-13, August 1978.
Oak Ridge National Laboratory, Report No. ORNL/TM-7581, April 1981.

6.1.2 **Operating, Maintenance and Production Costs** (cont'd)

Property taxes and insurance are assessed at 2.75% of the total plant installed cost (TPIC) comprising 2.0% local taxes and 0.75% insurance.

Plant overhead is assumed to be 40% of the operating and supervisory labour, maintenance and operating supplies. This cost includes plant management, engineering, fire protection, medical services, personnel, accounting, clerical and janitorial services.

Cost of coal, catalyst and chemicals, utilities, and waste disposal is treated as a variable expense. Cost of coal (Willow Bunch lignite) supplied to the plant is $9.75/tonne. This price was provided by Saskoil* and includes $0.28/tonne transportation costs from the mine to the plant.

Raw water is the only utility purchased from outside source at a cost of $1.10 per 1000 imperial gallons**. This cost has been estimated by SNC-Keith-Gabriel, Saskatchewan, and includes canal and pumping costs from Diefenbaker Lake to the plant site (see Appendix D-8).

Waste disposal cost (for ash) is assumed to be $2/tonne.

Running royalties are payment on an annual or semi-annual basis to the process licenser for the use of a proprietary process. The nature of the royalty payment is complex and requires more

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* F. Mourits, Coal Liquefaction Study Management, Letter to J.A. Mikhlin, Dec. 17, 1982.

** 1 imp. gallon = 4.546 litres.

6.1.2 **Operating, Maintenance and Production Costs** (cont'd)

detailed information than that possible within the scope of this assessment. Therefore, the running royalties have been neglected.

By product credits for sulfur at \$65/t, NH_3 at \$300/t and mixed phenols at \$250/t have been taken consideration. As explained earlier (Chapter 2.6), no credit for CO_2 have been taken.

Levelized capital and investment related charges for 30 years of operation are also presented in Table 6-2. In the first and second year of operation, the plant is operating at 50% and 90% of design capacity. To account for deficits in these two years, bridge financing at 10% interest rate in "real" terms and amortized over the balance period (28 years) is assumed.

6.1.3 **Product Cost**

Product costs for each of the four "preselected" coal liquefaction processes are presented on "Energy Value" and "Product Value" bases for relative ranking and selection of a process for Willow Bunch lignite. The product costs calculated are not intended for comparision with existing market prices.

The first approach, product cost on an "Energy Value" basis, is simply the unit energy cost ($\$/10^6$kJ) and is obtained by dividing the total annual costs by the total energy produced. The method is widely used but has the drawback that it does not consider the inherent quality and market preference of the individual products. For example, gasoline has a higher market value than fuel oil and calculating the product cost solely on heating value

6.1.3 **Product Cost** (cont'd)

would invert the price ranking of these fuels since heating oil has a higher fuel value per unit volume than does gasoline.

The second approach, product cost on value basis, considers both heating value and quality of the product. This method was developed by the Engineering Society Commission on Energy Inc. (ESCOE)*. In the ESCOE methodology, a value factor F_i is defined as the ratio of market value of individual product M_i to the chosen reference product M_r, thus

$$F_i = \frac{M_i}{M_r} = \frac{P_i}{P_r} \qquad (1)$$

where, M_i = market value of individual product
M_r = market value of the reference product
Pi = cost of individual product, \$/10^6 kJ
Pr = cost of reference product, \$/10^6 kJ

On an annual basis, a simplified overall product cost equation is

$$\sum P_i B_i = F + V + kC \qquad (2)$$

where, P_i = cost of individual product, \$/10^6 kJ
B_i = annual production (energy output) of product "i" in 10^6 kJ
F = annual fixed cost
V = annual variable cost

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* K.A. Rogers and R.F. Hill, "Coal Conversion Comparison", DOE Report No.FE-2468-51, July 1979.

6.1.3 **Product Cost** (cont'd)

k = levelized capital cost factor to allow for interest, taxes, return and recovery of capital investment, etc.
C = total capital investment.

Substituting the value of P_i from Equation (1) into Equation (2) gives:

$$P_r = \frac{F + V + kC}{\sum F_i B_i} \qquad (3)$$

Equation (3) is the basic cost equation relating the cost of a reference product P_r as a function of annual costs, annual production and the relative market values of various products. Thus, when P_r is determined, the cost of individual products on a "Value Basis" is obtained from Equation (1).

To determine product cost P_r, a set of market values M_i is required to calculate the value factors F_i. The market values of various products supplied by Petro-Canada are listed in Table 6-3. Value factors with reference to premium gasoline are also shown. The market values are wholesale price and includes Federal Excise Tax. Provincial sales tax is excluded.

Assigning the value of F_i as 1.0 in Equation (3) gives the product cost on "Energy Value" basis

$$P_e = \frac{F + V + kC}{\sum B_i} \qquad (4)$$

Table 6-3 Value Factor of Products from the Indirect Coal Liquefaction Plants

[Reference to Premium Gasoline in Saskatchewan]

Products	Market Value *	Converted Market ** Value in $\$/10^6$ kJ	Value Factor F_i
Fuel Gas	-	-	0.35 (assumed)
Propane	$14.80/bbl	4.82	0.40
Propane/Butane (LPG)	$19.00/bbl	5.01	0.42
Butane	$23.24/bbl	5.98	0.58
No.2 Fuel Oil	$29.30/bbl	7.34	0.61
Jet Fuel	$35.62/bbl	7.33	0.61
Diesel Fuel	$0.311/L	9.91	0.82
Unleaded Gasoline	$0.344/L	11.54	0.96
Mixed Alcohols	-	-	0.96 (assumed)
Methanol	$40.87/bbl	13.81	1.28
Premium Gasoline (Reference Product)	$0.359/L	12.04	1.00

* Market Value include Federal Excise Tax but Provincial Sales Tax is excluded. Prices for diesel, unleaded gasoline, and premium gasoline are wholesale prices for Saskatoon, Saskatchewan.

** Market value supplied by PETRO-CANADA has been converted to a consisted unit basis ($\$/10^6$ kJ) using conversion data given in ESCOE Report, FE-2468-51, July 1979.

Source:

M. Taherzadeh, Coal and Alternate Fuels, PETRO-CANADA, Telex to J.A. Mikhlin, SNC Inc., 7 April 1982

K.A. Rogers and R.F. Hill, ESCOE, DOE Report FE-2468-51, July 1979.

Table 6-4 Production Rate and Energy Output from the Coal Liquefaction Plants
[Capacity: 25,000 t/d of Willow Bunch Lignite]

[310 Operating days per year]

Coal Based Plant	Products	Flow rates		Heating Value	Energy Output, 10^{12} kJ/year	
		t/d	bbl/d	kJ/kg	$\sum B_i$	$\sum F_i B_i$
LSE Process	Propane/Butane (LPG)	389	4 157	45,100	5.43	2.28
- 65%	Gasoline	1 490	10,296	40,120	18.54	17.80
extraction	Jet Fuel	1,402	9,365	41,045	17.82	10.87
MAF basis	Diesel	1,262	8,102	41,200	16.12	13.22
					57.91	44.17
- 82%	Propane/Butane (LPG)	410	4,782	45,100	5.74	2.41
extraction	Gasoline	1,572	10,863	40,120	19.53	18.75
MAF basis	Jet Fuel	1,481	9,893	41,045	18.85	11.50
	Diesel	1,334	8,564	41,200	17.05	13.98
					61.17	46.64
Mobil	Fuel Gas	64	-	37,643	7.46	2.61
MTG	Propane	117	1,456	50,009	1.81	0.72
Process	Butane	186	2,062	48,940	2.83	1.64
	Gasoline	2,245	19,285	46,276	32.20	30.91
					44.30	35.88
	Fuel Gas	784	-	62,460	15.16	5.31
SASOL Type	Propane	69	855	49,923	1.05	0.40
Plant via	Butane	11	117	50,395	0.16	0.09
F-T	Gasoline	1,055	9,340	46,915	15.35	14.74
Plant	Diesel	214	1,785	47,069	3.13	2.57
	Heavy Fuel Oil	63	483	46,206	0.90	0.55
	Alcohols	133	1,047	31,187	1.30	1.24
					37.05	24.91
Methanol Plant	Pure Methanol	5,714	43,761	22,670	40.19	51.44

6.1.3 **Product Cost** (cont'd)

Table 6-4 presents the production rate and annual energy output for the preselected coal liquefaction processes. Using the annual, operating, and maintenance cost data given in Table 6-2, the levelized product costs on "Product Value" basis (Equation 3) and "Energy Value" basis (Equation 4) are calculated and presented in Table 6-5. Individual product costs are also presented.

The breakdown of the product cost into three main components (coal cost, operating and maintenance cost, and capital cost) is shown in Table 6-6.

Table 6-6 Components of Product Cost

	LSE	Mobil-MTG	SASOL-Type F-T Plants	Methanol Plant
Coal Cost, %	13.8	15.1	13.4	16.9
Operating and Maintenance Cost, %	32.6	34.1	34.9	33.8
Capital Cost, %	53.6	50.8	51.7	49.3

Table 6-7 Comparison of Preselected Coal Liquefaction Processes

	LSE				
	65% extraction MAF basis	82% extraction MAF basis	Mobil MTG	SASOL-Type F-T Plant	Methanol Plant
Desired Liquid Product * bbl/d	27,763	29,320	19,285	11,608	43,761
Desired Liquid Product Yield					
bbl/t as-received coal to plant	1.11	1.17	0.77	0.46	1.75
bbl/t MF coal to plant	1.71	1.80	1.19	0.71	2.69
bbl/t MAF coal to plant	2.16	2.27	1.50	0.91	3.41
Total Product Yield **					
bbl/t as-received coal to plant	1.12	1.19	0.94	0.94	1.75
bbl/t MF coal to plant	1.73	1.83	1.45	1.44	2.69
bbl/t MAF coal to plant	2.19	2.32	1.84	1.82	3.41
Product Efficiency					
Overall	57.5	60.7	43.5	36.4	39.5
Total Liquid	51.0	54.0	31.4	18.9	39.1
Desired Liquid *	51.0	54.0	31.4	18.0	-
Make-up Water Consumption, L/s	660	637	465	564	462
Capital Investment, $ million	2,639	2,639	2,213	2,545	1,926
Investment, $/daily bbl of total products **	93,740	88,765	93,770	109,035	44,010
Investment, $/daily bbl desired liquid	95,054	90,000	114,750	219,245	44,010

* Desired liquid products for direct and indirect process refer to gasoline, jet fuel, diesel

** Total product refers to all the products given in Table 6-4.

6.2 PROCESS COMPARISON AND SELECTION

In the previous section, a comparative techno-economic data were presented for the preselected coal liquefaction processes.

Analyzing the product output and distribution shown in Table 6-4, it may be seen that a wide range of products in varying amounts is produced in different processes. Comparison of total amount of desired liquid products (gasoline, jet fuel, diesel and fuel oil), show that the LSE process has the highest yield (4154 to 4387 t/d) followed by Mobil-MTG (2245 t/d) and finally the lowest yield in the by SASOL-type F-T plant (1332 t/d). However, all the liquid produced in Mobil-MTG is gasoline while in the LSE process 36 wt% is gasoline, 34 wt% jet fuel and the remaining 30 wt% diesel fuel.

Other pertinent data are presented in Table 6-7. As above, the total product yield is highest* for the LSE process followed by Mobil-MTG and the lowest for SASOL-type F-T plant. Similar trends are observed in overall and liquid product efficiencies. However, make-up water requirement is lowest for Mobil-MTG plant.

The total investment for a 25,000-t/d plant is about the same for the LSE and SASOL-type F-T plants and is higher than for Mobil-MTG. However, investment per daily production of desired products is lowest for LSE, followed by Mobil-MTG and highest for F-T plant. In other words, a plant based on equal output of products would have the lowest capital investment for the LSE process and highest capital investment for F-T plant.

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* The product yield from methanol plant are not compared with other processes because of different nature of the product. The performance data of methanol is discussed separately.

In summary, the SASOL-type F-T plant has the highest capital investment and product cost, lowest product yield and efficiency but at present the process is the only commercially proven coal liquefaction process. The Mobil-MTG and LSE processes appear to be comparable. Mobil-MTG has lower product yield and efficiency than the LSE process but has a better commercialization prospect.

In the case of methanol plant, the product yield is 2.69 bbl/t MF coal whereas the product efficiency is relatively low (about 39%) primarily due to the indirect method of liquefaction. Advantages are lower capital cost, much higher production and the product costs are comparable to present-day market prices (which are based on use of methanol as chemical feedstock only). The disadvantages are that it has lower energy density and there are numbers of practical problems associated with its use as fuel, such as uncertain market demand, distribution and storage difficulties, cold starting problems, lower tolerance for water, etc.

The pertinent properties of gasoline produced in all three process alternatives are shown in Table 6-8 in comparison with ASTM standard specifications (D-439). The products have varying characteristics but are of good quality and meet most of the specifications. Although the content of troublesome durene specifically present in MTG gasoline is below the target specifications (5 wt%), it would be preferable if it were below 4.0 wt% because it causes carburetor icing and engine stalling in low temperature conditions. Reduction of durene content may be achieved by blending with petroleum-derived gasoline.

Table 6-8 Comparison of Unleaded Gasoline Specification with Gasoline Properties of Preselected Coal Liquefaction Processes

	ASTM D-439	LSE Process	Mobil-MTG	SASOL-Type F-T Plant
API Gravity	-	52	61	67
Reid Vapor Pressure, kPa	105 max.	83	62	69
Distillation Analysis, °C				
IBP	-	32	30	30
10 vol%	55 - 65	65	44	42
50 vol%	76 - 115	109	93	86
90 vol%	185 max	183	169	168
FBP	225 max	208	198	215
Octane Rating				
Research Octane Number	87 min.	98	93	91
Motor Octane Number	82 min.	89	83	83
Antiknock Index $\frac{\text{RON+ MON}}{2}$	93 - 95	93.5	88	87
Copper Strip Corrosion	No.1 max.	No.2	N/A	N/A
Sulfur Content, ppm	1500 max	0.5	NIL	NIL
Leads Content	0.05 gm/US gal	NIL	NIL	NIL
Durene Content, wt%	5 max	N/A*	4.6	N/A*

* Not available, but usually much below the target specification

Sources:

W.P. Syrett, NCB, Personnel Communication to J.A. Mikhlin, SNC Inc., NCB Reference No. DR 496/2, 7/3, Jan. 25, 1982

M. Schreiner, Mobil Research and Development Corporation, DOE Report No. FE-2447-9, Jan. 1978

The jet fuel and diesel properties with standard specifications are shown in Table 6-9. The jet fuel from the LSE process has higher density than the specified value but it can be controlled by adjusting the jet-fuel cut. The diesel properties of both LSE and SASOL-type F-T plant meets the standard specifications.

The heavy oil produced only in SASOL-type F-T plant may be used as turbine fuel oil. The mixed alcohol byproduct is free from acids, aldehydes, ketones and water.

Methanol produced in the methanol plant is 99.95 wt% pure. The impurities are 0.05 wt% water and traces of other alcohols. The methanol meets the gasoline blendstock specifications ($<$ 1 wt% water and impurities).

Table 6-9 Comparison of Jet Fuel and Diesel Specifications with Properties of Products Produced in the Preselected Coal Liquefaction Processes

	Standard Specification	LSE Process	SASOL-Type F-T Plant
	ASTM D-1655		
Jet Fuel			
API Gravity	37 - 51	33.8	
Specific Gravity	0.7753 - 0.8398	0.856	
Distillation Temperature, °C			
10%	205 max.	184	
EP	311 max.	199	
Net Heat of Combustion, kJ/kg	42,745	43,682	
	ASTM D-975		
Diesel Fuel			
API Gravity	-		56
Distillation Temperature, °C			
90 vol%	287 max.	277	N/A
Cetane Number	40 min.	48	50
Flash Point, °C	38 min.	92	52
Corrosion, max.	No. 3 max.	N/A	No.1
Sulfur, ppm	5000	180	-
Ash, wt%	NIL	0.01 max.	traces

Sources:

W.P. Syrett, NCB, Personnel Communication to J.A. Mikhlin, SNC Inc., NCB Reference No. DR 496/2, 7/3, Jan. 25, 1982

M. Schreiner, Mobil Research and Development Corporation, DOE Report No. Fe-2447-9, Jan. 1978

7.0 CONCLUSIONS AND RECOMMENDATIONS

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7.0 CONCLUSIONS AND RECOMMENDATIONS

Review and evaluation of various direct and indirect coal liquefaction and pyrolysis processes were carried out for production of liquid fuels from designated Saskatchewan lignites in accordance with the main objectives of the study. Product yields and distribution, product quality, process status and suitability of Saskatchewan lignites were also examined. Processes selected in the prescreening evaluation (Stage I of the Study) were considered further for a more detailed technical and economic assessment (Stage II of the Study). Willow Bunch lignite was designated as a coal feedstock for techno-economic assessment of the preselected processes.

On the basis of preliminary evaluation and prescreening, the following conclusions and recommendations may be drawn.

- High-ash and moisture content of designated Saskatchewan lignites have a significant impact on product yields, efficiencies, and ranking of the processes.

- For direct coal liquefaction, a two-step approach comprising a coal "extraction" stage followed by selective hydrogenation of coal-derived liquid after removal of ash and residue seems to have an inherent advantage for processing of high-ash content coals. The LSE process having the two-step approach ranked as number one and was selected for techno-economic assessment of Willow Bunch lignite.

- Two other direct coal liquefaction processes, H-Coal and EDS, could be suitable for liquefaction of Saskatchewan lignites, if the high-ash content can be economically reduced to about 10 wt% MF coal or lower. Pretreatment/beneficiation of as-received lignites to reduce ash content was beyond the scope of study and investigation on this matter is recommended for future work.

- The development of the EDS process in the "bottoms recycle" mode may have a potential benefit for high-ash content lignites because of possible catalytic effect of the recycled ash and increased liquid product yields. The experimental evaluation of this mode was recently completed in an EDS pilot plant with Texas lignite. The results of these tests are not available yet but this aspect should be evaluated in future work.

- For indirect coal liquefaction, entrained-flow gasifiers are attractive for syngas production because of negligible methane content and the ability of the gasifier to handle fines. At present, the first-generation Koppers-Totzek entrained-flow, low-pressure gasifier appears to be most suitable for Saskatchewan lignites. This gasifier is commercially proven on lignites with ash content up to 32 wt% MF coal.

- The second-generation fluid-bed pressurized gasifiers, high-temperature Winkler (HTW) and Westinghouse, appears to be next best for gasification of Saskatchewan lignites. Experimental tests required by process developers to confirm the suitability of these gasifiers were beyond the scope of the study. In future work, techno-economic evaluation of these pressurized fluid-bed gasifiers, based on coal gasification tests, should be carried out in comparison with the entrained-flow Koppers-Totzek gasifier.

- For pyrolysis of Saskatchewan lignites, the Lurgi-Ruhrgas (L-R) process may be viable for Estevan lignite if a limited amount of liquid products is acceptable and surplus char can be marketed or utilized effectively. For Willow Bunch lignite, the L-R process is not viable from a liquefaction point of view because of very low liquid product yields. The L-R process was not considered for techno-economic assessment (Stage II of the Study).

As a result of techno-economic assessment of preselected processes (Stage II of the Study), the conclusions and recommendations are as follows:

- SASOL-type Fischer-Tropsch synthesis plant in combination with Koppers-Totzek gasifier configuration has lower product yields and efficiencies as well as high investment and product cost in comparison to Mobil-MTG or LSE process. On the other hand, the SASOL-type plants are the only commercially proven coal liquefaction operations, at present.

- The LSE process and the Mobil-MTG are comparable. The LSE process has a higher estimated liquid product yield (2.27 bbl/t versus 1.5 bbl/t MAF coal) and higher liquid product efficiency (54% versus 31.4%). On the other hand, in the LSE process, only 34 wt% of liquid product is gasoline (the balance 36 wt% is jet fuel and 34 wt% is diesel) while 100% of liquid yield in Mobil-MTG is gasoline. Product costs are nearly the same for LSE and Mobil-MTG processes.

- Mobil-MTG has better commercialization prospects than the LSE process since the only part of the Mobil process which has not been demonstrated commercially is the conversion of methanol to gasoline.

- The capital investment for a 25,000-t/d plant processing Willow Bunch lignite is lower for Mobil-MTG (Can. $2,213 million) compared to LSE process (Can. $2,639 million). However, the investment per daily production of transportation fuels is lower for LSE process ($90,000/bbl versus $114,750/bbl). If, however, gasoline is the only desired product, then the invesment per daily barrel for Mobil-MTG would be almost half that of LSE process.

- Detailed conceptual design of a commercial plant with economic and financial analyses (including sensitivity cases) is recommended for both LSE and Mobil-MTG processes in order to further distinguish their merits. A market survey to determine the demand for transportation fuels in the Province of Saskatchewan is also required.

- If the market survey will show that methanol is a required product, the coal-based plant for the production of pure methanol appears attractive. All the processing areas comprising such a plant has been demonstrated on a commercial scale.

In addition to assessment of advanced direct and indirect coal liquefaction and pyrolysis processes, a preliminary review of new processes or concepts considered for commercialization around 1995 was carried out. The list of processes evaluated, their present demonstration scale, R&D status (including funding status) are summarized in Table 7-1. Details of the evaluaton are presented in Volume IV. Conclusions and recommendations are given below:

- Amongst the direct coal liquefaction, two processes (CO-Steam and SRL) have been investigated specifically for lignites. Other direct coal liquefaction processes have not been tested on lignites but some characteristic processing steps/features are attractive for coal liquefaction, including lignites such as the use of emulsified catalyst (as in Dow process) and solvent deashing approach (as in TSL and SRC processes).

- The coprocessing approach has a number of potential advantages especially for the Province of Saskatchewan where lignite and heavy oil resources are located in close proximity. However, uptil now, the coprocessing concept has been investigated mostly on autoclave level and many problem areas have yet to be resolved on bench-scale and PDU levels before demonstration on a pilot-plant level and further commercialization of the process.

- The Occidental pyrolysis process may be attractive for Estevan lignite if in addition to liquid product, all the char can be utilized effectively.

- The process development work in most of the processes has been discontinued due to financial constraints and present oil glut situation in the world. Some of the processes have fundings till the end of 1982.

7.0 CONCLUSIONS AND RECOMMENDATIONS (cont'd)

In addition, an independent experimental program was also carried out concurrently with this study by the Chemical Technology Division, B.C. Research, and the results are presented in a separate report entitled, "Liquefaction and Pyrolysis Tests on Samples of Estevan and Coronach Lignites", Project Report 2-02-691, March 1982.

Samples of Estevan and Willow Bunch (Coronach) lignites were tested to determine the conversion products obtained by pyrolysis using the Modified Fischer-Schrader Assay test and by direct hydroliquefaction using a tubing bomb experimental unit. An Australian Morwell lignite was also tested in a tubing bomb unit for comparative liquefaction analysis. A summary of the results is given in Table 7-1 and the conclusions are presented below:

- Pyrolysis tests indicated low tar yields for Willow Bunch lignite ($\simeq$ 6-7 wt% MAF coal) and somewhat higher tar yields for Estevan lignite ($\simeq$ 9.5-10.5 wt% MAF coal). Char yields for both lignites are high (64-66 wt% MAF coal).

- Liquefaction test results indicate wide variation in product yields depending on experimental conditions.

- Total liquid yields (i.e. THF solubles including chemically produced water) from Estevan lignite varied from 16 wt% MAF coal at 460°C under non-catalytic condition to 73.4 wt% MAF coal at 440°C with HDS-1442A Co/Mo catalyst. Corresponding liquid yield ranges for the Willow Bunch and Australian Morwell lignite were 23-71.9 wt% MAF coal and 10.7-78.4 wt% MAF coal, respectively. On this basis, B.C. Research ranking of the coal is Morwell $>$ Estevan $\geq$ Willow Bunch.

Table 7-1 Process Demonstration Scale and Status of New Processes or Concepts Considered for Commercialization around 1995

Process	Demonstration Scale	R&D Status	Funding Status
Direct Coal Liquefaction (One-Stage Hydrogenation)			
- Dow	0.1 t/d	In progress	Current (till Dec. 82)
- CO-Steam	2.3 kg/h (Dismantled)	Abandoned	None
Direct Coal Liquefaction (Two-Stage Hydrogenation)			
- Consol Synthetic Fuel (CSF)	9.0 t/d (Dismantled)	Abandoned	None
- Two-Stage Liquefaction (TSL)	0.5 t/d	In Progress	Current (till Dec. 82)
- Solvent-Refined Coal (SRC)	45 t/d	Demo. Plant Eng. Design In Progress	Current (till Dec. 82)
- Zinc Halide (Z-H)	1.0 t/d	Abandoned	None
- Solvent-Refined Lignite (SRL)	0.5 t/d Dismantled	Abandoned	None
Indirect Coal Liquefaction (Gasification Processes)			
- Shell	150 t/d	In Progress	Not Determined
- U-Gas	25 t/d	In Progress	Not Determined
Coal-Oil Coprocessing			
- Coil	30 bbl/d	Abandoned	None
- Japanese Solvolysis Coal Liquefaction (JSCL)	1.0 t/d	In Progress	Not Determined
- Veba Coal/Oil Process	auto-clave	In Progress	Not Determined
Pyrolysis			
- Occidental	2.5 t/d	Abandoned	None

7.0 CONCLUSIONS AND RECOMMENDATIONS (cont'd)

- In terms of distillate liquid product (< 525°C), under the best yield conditions (440°C, catalyst), the range of observed liquid yields (57.0-62.6 wt% MAF coal) are nearly the same for all the three lignites, as they are within the estimated limits of experimental error.

- Finally, the experimental conditions used in the experiments are too specific for comparison with other coals.

REFERENCE AND SOURCES

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REFERENCES AND SOURCES

General

H.I. Smith, "Coal Conversion Technology", Chemical Technology Review No. 66, Noyes Data Corporation, N. J., 1976.

J.D. Batchelor, U.S. Department of Energy, "Technical Status of the Large Coal Liquefaction Programs at the Department of Energy", paper presented at Institute of Gas Technology sponsored Symposium on Advances of Coal Utilization Technology Louiseville, Kentucky, May 14-18, 1979.

N. Berkowitz, "Introduction to Coal Technology", Academic Press, N. Y., 1979.

P. Nowacki, "Coal Liquefaction Processes", Chemical Technology Review No. 131 , Energy Technology Review No. 45, published by Noyes Data Corp. N. J., 1979.

P. Sterbenz, "Coal Conversion Today, Hydrocarbon Processing, p. 68-72, July 1980.

D. J. Deutsch, "Solid/Liquid Separation Routes are Nearing Demonstration Stage", Chemical Engineering, p. 33-35. August 25, 1980.

B.R. Rodger et al., Oakridge National Laboratory, "Assessment of Current Research and Development in Support of the U.S. Coal Liquefaction Demonstration Plants Program", paper presented at the 15th Intersociety Energy Conversion Engineering Conference, Seattle, Washington, August 18-22, 1980, Proceedings Vol. 3, p. 1812-1817.

L.M. Joseph, U.S. Department of Energy, "The Department of Energys' Major Project Coal Liquefaction Program". Paper presented at the 15th Intersociety Energy Conversion Engineering Conference, Seattle, Washington, August 18-22, 1980, Proceedings Vol. 3, p. 1808-1811.

D. Oliver, Financial Time Business Information Ltd., "Oil and Gas from Coal", Financial Times Business Information Ltd., Bracken House, London, January 1981.

Overview of Direct Coal Liquefaction Processes

H-Coal Process

A.L. Conn et al, American Oil Company, Research and Development Department, "Evaluation of Project H-Coal", prepared for Department of Interior, Office of Coal Research, under Contract No. 14-01-0001-1188, December 1, 1967.

H.H. Stotler, Hydrocarbon Research Inc., "The H-Coal Pilot Plant Program", Chemical Engineering Progress Technical Manual on Coal Processing Technology, Vol. 2, published by American Institute of Chemical Engineers, 1975.

REFERENCES AND SOURCES (cont'd)

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant", Work Sponsored by Electric Power Research Institute under Contract EX-76-C-01-1544, EPRI Report No. AF-681, Project No. 238-1, March 1978.

W. Karolkiewicz, Hydrocarbon Research Inc., "Solids-Liquid Separation in H-Coal Development", paper presented at AIChE meeting, Philadelphia, Pennsylvania, June 1978.

Hydrocarbon Research Inc., "H-Coal Run Summary of Bench-Scale and Process Development Unit Operations", Issued by HRI, R & D Centre N.J., October 14, 1979.

Ashland Synthetic Fuels, Inc., "H-Coal Pilot Plant, Cattlettsburg, Fact Sheet", Information Sheet published by Ashland Oil Inc., November 1979.

P.A. Buckhingham, R. Ravikumar, and A.A. Leavitt, Fluor Engineers and Constructor Inc., "Engineering Evaluation of Conceptual Coal Conversion Plant Using the H-Coal Liquefaction Process", prepared for Electric Power Research Institute under Contract No. RP-411-4, Report No. AF-1297, December 1979.

H.H. Stotler and R.T. Shutter, "Status and Plans of H-Coal Pilot Plant", Chemical Engineering Progress Technical Manual on Coal Processing Technology, Vol. 5, published by American Institute of Chemical Engineers, 1979.

The Energy Daily, National Press Building, Washington, Thursday, February 7, 1980, Thursday July 31, 1980.

J.A. Mikhlin, SNC Inc., Visit to Hydrocarbon Research Inc., H-Coal Experimental Facilities and Laboratories, Lawrenceville, N.J. March 1980.

J.A. Mikhlin, SNC Inc., Visit to H-Coal Pilot Plant at Cattlettsburg, Kentucky, and meeting at HRI Inc., Lawrenceville, N.J., August 11-14, 1980.

J.L. Gendler, HRI, Personal Communication to Dr. J.A. Mikhlin, SNC Inc., HRI Ref. No. 01-JLG-17-81, January 23, 1981.

HRI Report, "H-Coal Evaluation for the Energy Authority of New South Wales", August 1981.

R.M. Eccles and G.R. DeVaux, Hydrocarbon Research Inc., "Current Status and Commercialization", paper presented at the 2nd World Congress of Chemical Engineering of Chemical Engineering held in Montreal, Canada, October 4-9, 1981. Proceeding Vol. 2. pp. 514-520.

The Energy Daily, National Press Building, Washington, "H-Coal Lays Plans for When the DOE Money Runs Out", January 19, 1982.

REFERENCES AND SOURCES (cont'd)

RAG/Veba Process

H.E. Wurfel, Saarberfweker AG, "I.G. Farben Process Further Development of a Technically Proven Process", paper presented at the Symposium of Coal Refining, Edmonton, Alberta, April 20-21, 1978.

B. Strobel, I. Romey and G. Kolling, Bergbau-Forshung GmbH, "German Routes to Coal Hydrogenation", paper presented at Institute of Gas Technology sponsored Symposium on Advances in Coal Utilization Technology at Louiseville, Kentucky, May 14-18, 1979.

Birtley Coal and Mineral Testing, and Cascade Coal Petrography Ltd., "Proximate, Ultimate and Petrographic Analysis of Canadian Subbituminous Coal from Forestburg, Alberta", analysis done for SNC Inc., Montreal, October 1979.

G. Escher et al., "Ruhrkohle/Veba Oel Technology for Hydrogenation of Hard Coal", paper presented at Annual Fall Conference on the Canadian Society for Chemical Engineering, Edmonton, Alberta, October 19-22, 1980.

J.A. Mikhlin, SNC Inc., Report of the Ruhrkohle AG/Veba Oel AG/SNC meeting on February 16, 1981, Essen, Federal Republic of Germany.

J.A. Mikhlin, SNC Inc., visit to Beregbau-Forschung coal liquefaction facilities, Essen, and 200 tonne-per-day pilot plant at Bottrop, Federal Republic of Germany, February 16-17, 1981.

Veba Oel AG, Buer, Federal Republic of Germany, "Description and Flowsheet of the Bottrop Pilot Plant", February 20, 1981.

Veba Oel AG, Buer, Federal Republic of Germany, Technical Information on "Conceptual Design Data of Commercial Plant for Coal Hydrogenation" February 20, 1981.

Veba Oel AG, Buer, Federal Republic of Germany, Technical Information on "Upgrading of Oil from Coal", February 20, 1981.

Veba Oel AG, Buer, Federal Republic of Germany, Information Package and Report on "Autoclave Tests of the Liquefaction Behaviour of Canadian Subbituminous Coal", July 29, 1981.

J.A. Mikhlin, SNC Inc., Visit to Veba Oel AG, October 29-November 2, 1981.

Rheinbaun Process

D. Boecker et al., Rheinische Braunkohlenwerke AG and Union Rheinische Braunkohlen Kraftstoff AG, "Rheinbraun Liquefaction of Brown Coal", paper presented at EPRI/KFA- Julich Conference on Synthetic-Fuels: Status and Direction, at San Francisco, October 1980.

REFERENCES AND SOURCES (cont'd)

J.A. Mikhlin, SNC Inc., Visit to Rheinische Braunkohlenwerke AG, Koln, February 19, 1981; and October 29 - November 2, 1981.

D. Boecker, W. Dolkemeyer, K.-H. Keim, U. Lenz and G. Ritter, "Results of Liquefaction of Rheinish Brown Coal", paper presented at the "Lignite Symposium", held at San Antonio, June 1981.

EDS Process

ER and E Co., "EDS Commercial Plant Study Design Update Offsite Design Basis - Illinois Coal Market Flexibility Sensitivity Case", prepared for U.S. Department of Energy under the contract No. EF-77-A-01-2893, Report No. FE-2893-37, July 1979.

D.F. Ryon, B.H. White, and S. Zaczepinski, Exxon Research and Engineering Co. "Characterization and Processibility of EDS Product Streams, Proceedings of 5th EPRI Contractor's Conference on Coal Liquefaction, Palo Alto, CA, May 1980.

A.S. Crocker "EDS Commercial Plant Study Design Update Comparison of Base Coal with Market Sensibility Case Economics Sensitivities", Synthetic Fuel Research, Exxon Research and Engineering Company, Florham Park, N.J. August 7, 1980.

J.A. Mikhlin, Visit to Exxon Research and Engineering Company and EDS Experimental Facilities at Florham Park, N.J. and Pilot Plants, Baytown, Texas, August 27028, 1980.

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal, Volume 2 - Base Case", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-61, Vol. 2, March 1981.

W.R. Epperly, K.W. Plumplee, and D.T. Wade, "Exxon Donor Solvent Coal Liquefaction Process: Development Program Status VI", Chemical Engineering Progress, May 1981.

S.J. Hsia, K.L. Trachte, G.A. Wilcox and R.D. Williams, "Liquefaction of Low Rank Coals with the Exxon Donor Solvent Coal Liquefaction Process", presented at the 11th Biennal Lignite Symposium, San Antonio, Texas, June 15-17,1981.

A.M. Goldsteim, "The Exxon Donor Solvents Process: Equipment and Scale-up Issues", paper presented at the Second World Congress of Chemical Engineering, Montreal, Canada, October 4-9, 1981.

J.W. Taunton, K.L. Trachte, and R.D. Williams, Exxon Research and Engineering Company, "Coal Feed Flexibility in the Exxon Donor Solvent Coal Liquefaction process" 1981.

REFERENCES AND SOURCES (cont'd)

LSE Process

J.C. Whitehead and D.F. Williams, National Coal Bord, J. Inst. Fuel. p. 182 (1971).

M.D. Gray, Coal Processing Consultants Ltd., "Liquids from Coal Processes Under Development by the National Coal Board", UK, paper presented at 29th Canadian Chemical Engineering Conference on Synthetic Liquids from Coal and Biomass held at Sarnia, Ontario, September 30 to October 3, 1979.

J.A. Mikhlin, SNC Inc., Visit to Experimental Facilities of Coal Research Establishment, National Coal Board, February 1981.

G.G. Thurlow, National Coal Board, "Producing Fuels and Chemical Feedstocks from Coal", Chem. Eng., Progr. pp. 81-84, March 1980.

National Coal Board, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", report prepared by NCB, Coal Research Establishment, Cheltenham, February 1981.

D.F. Williams, Head of Coal Refining, National Coal Board, Personal Communication to J.A. Mikhlin, SNC Inc., dated February 25, 1981.

W.P. Syrett, National Coal Board, Personal Communication to J.A. Mikhlin, SNC Inc., "The Processing of Saskatchewan Lignites by the NCB's Liquid Solvent Extraction Process", NCB Reference No. DR3000, DR496/2, October 1981.

Overview of Coal Pyrolysis

Lurgi/Ruhrgas Process

R.W. Rammler "The Retorting of Coal, Oil Shale and Tar-Sand by Means of Circulated Fine-Grained Heat Carriers as a Preliminary Stage in the Production of Synthetic Crude Oil", paper presented at the AIME Annual Meeting, Hydrocarbon Session, Denver, Colorado, Feb. 15-19, 1970.

R.W. Rammler, "Retorting of Coal, Oil Shale and Tar Sand by Means of the Lurgi-Ruhrgas Process", modified version of the paper presented at the German-Canadian Coal Refining Symposium, Edmonton, Alberta, April 20-21, 1978.

R.W. Rammler, "The Lurgi-Ruhrgas Process for Carbonization of Coal", paper presented at the EPRI Conference on Coal Pyrolysis, Palo alto, February 25-26, 1981.

B.C. Resources Laboratories, "Liquefaction and Pyrolysis of Estevan and Coronach Lignites", Project Report No. 2-02-691, March 1981.

REFERENCES AND SOURCES (cont'd)

Overview of Indirect Coal Liquefaction

F. Boegner, Davy International AG, T.J. Kendron, T.K. Subramanian, Davy McKee Corp., "Winkler Fluid Bed Coal Gasification Experience With Low Grade Coals", proceedings of Seventh Annual International Conference on Coal Gasification, Liquefaction and conversion to Electricity, held at University of Pittsburgh, Pittsburgh, PA, August 5-7, 1980, pp. 217.

W. Koog, Texaco Development Corporation, "The Texaco Coal Gasification Process- Status and Outlook", The Eight Annual WATTec Energy Conference and Exhibition, Knoxville, TN, February 18-20, 1981.

W. Adlhock and K.A. Theis, Rheinische Braunkohlenwerke AG, "The Rheinbraun High-Temperature Winkler (HTW) Process", EPRI Conference Proceedings: Synthetic Fuels-Status and Directions, Vol. 2: EPRI WS-79-238, May 1981, San Francisco, CA, pp 29.1

B. Cornils and R. Specks, Ruhrchemie AG, "Experiences with the Texaco Process of Coal-Dust Pressure Gasification Using the Ruhrchemie/Ruhrkohle Technical Version", EPRI Conference Proceedings: Synthetic Fuels Status and Directions, Vol. 2: EPRI WS-79-238, May 1981, San Francisco, CA, pp.19-1.

J.D. Holmgren, C.E. Seglem, L.A. Salvador, and M.W. Dyos, Westinghouse Electric Corporation, "The Westinghouse Gasification Program". EPRI Conference Proceedings: Synthetic Fuels - Status and Directions, Vol. 2: Section 18- Appendices EPRI WS-79-238, May 1981, SAN Francisco, CA, pp. 28-1.

C. Lohmann and J. Langhoff, Ruhrkohle AG, "Ruhr-100 Development Work on the Pressurized Lurgi Gasifier in Dorsten" EPRI Conference Proceedings: Synthetic Fuels: Status and Directions Vol. 2: WS-79-238, May 1981, San Francisco, CA, pp. 23-1.

P.E. Rudolph, C. Hafke, and P.K. Herberg, Lurgi Kohle Und Mineralöltechnik GmbH "Lurgi Coal Gasification", EPRI Conference proceedings: Synthetic Fuel- Status and Directions, Vol. 2: EPRI WS-79-238, May 1981, San Francisco, CA, pp. 22-1.

R.B. Sharman and J.E. Scott, British Gas Corp. "The British Gas/Lurgi Slagging Gasifier: Relevance to Power Generation". EPRI Conference Proceedings: Synthetic Fuel-Status and Directions Vol. 2, EPRI WS-79-238, May 1981, San-Francisco, CA, pp. 24-1.

F.B. Walter, H.C. Kaufman, Texaco Inc. and T.L. Reed, Southern California Edison Co. "The Cool Water Coal Gasification Combined Cycle Technology" EPRI Conference Proceedings: Synthetic Fuels-Status and Directions. Vol. 2: EPRI WS-79-238, May 1981, San-Francisco, CA, pp. 20-1.

REFERENCES AND SOURCES (cont'd)

J.A. Mikhlin, SNC Inc., Visit to Lurgi Mineralöltechnick GmbH, Lurgi Research and Development Centre, October 29, November 5, 1981.

J.A. Mikhlin, SNC Inc., Visits to Rheinbraun on HTW Coal Gasification, cologne, Ruhrchemie/Ruhrkohle/Texaco Coal Gasification, Oberhausen, British Gas Corporation, on British Gas/Lurgi Gasifier UK, October 29-November 5, 1981.

J.A. Mikhlin, SNC Inc., Visit to Westinghouse Research and Development Centre and Waltz Mills Pilot Plant on Westinghouse Coal Gasification Process, January 1982.

H. Staege, Gesellschaft für Kohle-Technologie mbH. "Entrained-Bed Gasification" Hydrocarbon Processing, March 1982, pp. 92-96.

J.A. Mikhlin, SNC Inc., Visit to Krupp-Koppers, Krupp Research and Development Centre, Essen, on Krupp-Koppers Coal Gasification process, April 1982.

W.B. Crouch, "The Texaco Coal Gasification Process - Synthesis Gas for Chemical Feedstocks", International Coal Conversion Conference, August 1982, Pretoria, South Africa.

M. Ilgner, K.H. Lim and E. Nitschke, "Considerations for the Design of a Commercial Coal Gasification Plant According to the Texaco Process", International Coal Conversion Conference August 1982, Pretoria, South Africa.

P.E. Rudolph and C.P. Hafke, "Lurgi Coal Gasification- Achievement and Developments", International Coal Conversion Conference, August 1982, Pretoria, South Africa.

L.A. Salvador and P.J. Meyer, "The Joint SASOL/Westinghouse Commercial Scale Fluidized Bed with a Draught Tube", International Coal Conversion Conference, August 1982, Pretoria, South Africa.

H. Teggers, W. Adlhoch, V. Femmer, and K.A. Theis, "Actual Status of the Rheinbraun High-Temperature Winkler (HTW) Gasification Process", International Coal Conversion Conference, August 1982, Pretoria, South Africa.

Techno-Economic Assessment of Preselected Processes

"Development Document for Effluent Guidelines and New Source Performance Standards for the Petroleum Refining Point Source Category", U.S. Environmental Protection Agency, EPA Report No. 440/1-74-014A, Washington D.C., April 1974.

Environment Saskatchewan, Air Pollution Control Regulations, 211/75, 1975.

REFERENCES AND SOURCES (cont'd)

Environment Saskatchewan, Water Pollution Control Branch, "Water Quality Objectives", January 1977, February 1977.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job No. C-8093, October 30, 1977.

B.C. McBeath, The Engineering Societies Commission on Energy Inc., "Guidelines for Economic Evaluation of Coal Conversion Processes, prepared for the U.S. Department of Energy under Contract No. EF-77-C-01-2468, Report No. FE-2468-44, April 1979.

K.A. Rogers and R.F. Hill, The Engineering Societies Commission on Energy Inc., "Coal Conversion Comparison, prepared for the U.S. Department of Energy Inc. under Contract No. EF-C-01-2468, Report No. FE-2468-51, July 1979.

M.D. Gray, Coal Processing Consultant Ltd., "Liquids from Coal Processes Under Development by the National Coal Board, U.K.", paper presented at 29th Canadian Chemical Engineering Conference on Synthetic Liquids from Coal and Biomass, held at Sarnia, Ontario, September 30 - October 3, 1979.

National Coal Board Report to SNC Inc., "Coal Liquefaction for Transportation Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", February 1981.

R.M. Wham et al, "Liquefaction Technology Assessment - Phase 1: Indirect Liquefaction of Coal to Methanol and Gasoline Using Available Technology", Oak Ridge National Laboratory Report ORNL-5664, February 1981.

K.A. Rogers, U.S. Synthetic Fuels Corp, "Synfuel Technologies and Costs", paper presented at the AIME Meeting, Chicago, Illinois, February 24, 1981, published in Material and Society, Vol.5, No.4, pp.355-363, 1981.

R. Salmon et al, Oak Ridge National Laboratory, "Technical and Economic Evaluation of Coal Liquefaction Technologies - Phase 0: Preliminary Screening Evaluation", Report No. ORNL/TM-7581, April 1981.

Construction Price Statistics, "Chemical and Mineral Process Plant Indexes", Table 19; and "Union Wage Rate Index", Table 3; November 1981, Catalogue No. 62-007, published by Statistics Canada, Ottawa.

F. Mourits, Coal Liquefaction Study Management, Saskoil, letter to J.A. Mikhlin, SNC Inc., December 17, 1981.

Engineering New Record, January 7, 1982, January 16, 1982.

REFERENCES AND SOURCES (cont'd)

W.P.Syrett, National Coal Board, report to SNC Inc. "The Processing of Saskatchewan Lignite by the NCB's Liquid Solvent Extraction Process: Willow Bunch Lignite", NCB Reference No., 496/2, 7/3, January 23, 1982. (The report is attached in confidential Volume V,"Process Developers Data and Information" of the present study).

G. Kimber, Coal Research Establishment National Coal Board, telex to J.A. Mikhlin, SNC Inc., February 22, 1982.

M. Taherzadeh, Coal and Alternate Fuels, PETRO-CANADA, Telex to J.A. Mikhlin, SNC Inc., 7 April 1982.

Construction Collective Agreement Summaries, "Labour Rate in Saskatchewan", published by the Canadian Construction Association, Ottawa, April 1982.

W.P. Syrett, National Coal Board, report to SNC Inc., "The Assessment of a Canadian Lignite for the National Coal Board Liquid Solvent Extraction Process", April 1982. (Report is attached in confidential Volume V "Process Developers Data and Information" of the present study).

H. Staege, Krupp Koppers GmbH, "Preliminary Study for the Production of Pure Methanol or Synthesis Gas from Coal in the Basis of the Koppers-Totzek Coal Gasification Process", Report to SNC Inc., Krupp Koppers Ref. No. VK-1/Mn-Ko-1813, June 1982. (The report is attached in the confidential Volume V "Process Developers Data and Information" of the present study).

APPENDIX A

TYPICAL ANALYSES AND PROPERTIES OF WILLOW BUNCH LIGNITE CONSIDERED FOR TECHNO-ECONOMIC ASSESSMENT OF PRESELECTED LIQUEFACTION PROCESSES

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Appendix A-1 Proximate Analysis and Gross Calorific Value of Willow Bunch Lignite (Coronach Coal)

	Average Coal Analysis (206 Samples)		
	As-Received	MF	MAF
Proximate Analysis, wt%			
Moisture	35.0	-	-
Volatile Matter	25.7	39.6	50.1
Fixed Carbon	25.0	38.5	48.8
Ash	13.7	21.0	-
Sulfur	0.6	0.9	1.1
Gross Calorific Value, kJ/kg	13260	20380	25848

	Worst Coal Analysis		
	As-Received	MF	MAF
Proximate Analysis, wt%			
Moisture	40.0	-	-
Volatile Matter	20.4	34.0	50.0
Fixed Carbon	19.9	33.2	48.8
Ash	19.2	32.0	-
Sulfur	0.6	0.8	1.2
Gross Calorific Value, kJ/kg	10476	17461	25995

	Best Coal Analysis		
	As-Received	MF	MAF
Proximate Analysis, wt%			
Moisture	30.0	-	-
Volatile Matter & Fixed Carbon	59.5	85.0	99.2
Ash	10.0	14.3	-
Sulfur	0.5	0.7	0.8
Gross Calorific Value, kJ/kg	15817	22594	26581

Sources: Letter from F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to J.A. Mikhlin, SNC Inc.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job. No. C-8093, October 30, 1977.

Appendix A-2 Ultimate and Petrographic Analysis of Willow Bunch Lignite (Coronach Coal)

	Average Coal Analysis (206 samples)	Worst Coal
Ultimate Analysis, wt% MF coal		
Carbon	52.1	45.0
Hydrogen	4.0	3.5
Nitrogen	0.5	0.3
Sulfur	0.9	0.8
Oxygen	21.5	18.4
Ash	21.0	32.0

Petrographic Analysis, %		Worst Sample	Best Sample
Huminite	Structured	45.6	74.0
	Groundmass	18.8	13.5
	Unstructured	3.2	2.3
Liptinite	Sporinite	1.0	-
	Resinite	-	-
	Cutinite	-	-
	Suberinite	-	-
	Liptodetrinite	1.6	1.0
Inertinite	Fusinite	12.6	1.7
	Semifusinite	5.4	3.7
	Sclerotinite	0.2	-
	Macrinite	0.8	-
	Inertodetrinite	8.8	0.7
Mineral Matter		2.0	3.1
		100.0	100.0

	Average (5 samples)
Huminite	77.6
Liptinite	1.7
Inertinite	18.7
Mineral Matter	2.0
	100.0

Sources: Letter from F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to J.A. Mikhlin, SNC Inc.

Petrographic Analysis from University of Regina Report to Saskoil, March 1982.

Appendix A-3: Ash Analysis of Willow Bunch Lignite (Coronach Coal)

	In Oxidizing Atmosphere (108 Samples)	In Reducing Atmosphere (69 Samples)
Ash Fusion Temperature, °C		
Initial Deformation	1153	1173
Softening Temperature	1211	1192
Hemispherical Temperature	1236	1199
Fluid Temperature	1293	1209

Chemical Analysis of the Ash	Average (133 Samples)
Silica, SiO_2	42.1
Alumina, Al_2O_3	15.6
Ferric Oxide, Fe_2O_3	6.7
Lime, CaO	14.3
Magnesia, MgO	5.0
Potassium Oxide, K_2O	1.3
Sodium Oxide, Na_2O	1.8
Titania, TiO_2	0.9
Balance	12.3

Ash Distribution No. of Samples	wt% MF coal
2	5 - 10
33	10 - 15
35	15 - 20
33	20 - 25
13	25 - 30
8	30 - 35
5	35 - 40
4	40

Sources: Letter from F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to J.A. Mikhlin, SNC Inc.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job. No. C-8093, October 30, 1977.

Appendix A-4: Size Analysis of Willow Bunch Lignite (Coronach Coal)

	Sample 1	Sample 2	Sample 3
Retained by screen 2"	51.9	35.5	27.1
Retained by screen 1½"	8.07	4.17	8.60
Retained by screen 1"	7.43	7.92	14.3
Retained by screen ½"	10.3	14.6	18.6
Retained by screen #3 (6.35 mm)	8.11	11.9	12.3
Retained by screen #8 (2.38 mm)	7.03	25.3*	9.02
Passing #8 (2.38 mm)	6.26		9.90
TOTAL	99.10	99.39	99.82

Sample 1: Coal from mine pit

Sample 2: Coal from conveyor belt immediately after primary crusher.

Sample 3: Coal from bottom of tressle (coal storage).

* This percentage will be broken in fractions retained and passing #8 sieve.

Sources: Letter from F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated december 15, 1981, to J.A. Mikhlin, SNC Inc.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job. No. C-8093, October 30, 1977.

APPENDIX B

ANALYSIS OF DIFFERENT RANK COALS TESTED IN ADVANCED COAL LIQUEFACTION AND PYROLYSIS PROCESSES

Table B-1 Analyses of Different Rank Coals Tested in Advanced Direct Coal Liquefaction and Pyrolysis Processes

Process	H-Coal		
Coal(s)	North Dakota Lignite[1]	Australian Brown, Muswellbrook[2]	Subbituminous Wyodak[3]
Moisture (as received), wt%	35.0	30.0	30.0
Gross Calorific Value (GCV) kJ/kg MF coal	25,335	23,260	26,563
Proximate Analysis, wt% MF coal			
Volatile Matter	44.0	30.6	44.6
Fixed Carbon	45.8	46.8	46.8
Ash	10.2	22.6	8.6
Ultimate Analysis, wt% MF coal			
Carbon	65.3	63.4	66.8
Hydrogen	4.5	3.7	5.0
Nitrogen	0.8	1.5	0.8
Sulfur	1.1	0.6	1.0
Oxygen	18.0	8.2	17.7
Ash	10.3	22.6	8.7
Petrographic Analysis, vol%			
Vitrinite		69.0	
Exinite	N/A	5.0	N/A
Inertinite		19.0	
Minerals Matter		2.0	
Ash Analysis, wt%			
Silica, SiO_2			29.3
Alumina, Al_2O3			18.8
Titania, TiO_2			1.1
Ferric Oxide, Fe_2O_3			6.6
Lime, CaO	N/A	N/A	18.8
Magnesia, MgO			4.5
Potassium Oxide, K_2O			0.4
Sodium Oxide, Na_2O			0.6
Phosphorous Pentoxide, P_2O_5			0.4
Sulfur Trioxide, SO_3			17.3
Undetermined			1.8

Notes/Sources:

1. Coal supplied by Peabody Coal Company from Dickinson Area, Stark County, North Dakota (HRI Report, Project H-Coal", 1967)
2. HRI Report, "H-Coal Evaluation for the Energy Authority of New South Wales", August 1981
3. EPRI Report No. AF-1143-SR, July 1979.

300

Table B-1 Analyses of Different Rank Coals Tested in Advanced Direct Coal Liquefaction and Pyrolysis Processes (cont'd)

Process / Coal(s)	H-Coal	RAG/Veba	
	Bituminous Illinois No. 6[1]	Bituminous German[2]	Subbituuminous Forestburg[3]
Moisture (as received), wt%	11.0	4.3	25.6
Gross Calorific Value (GCV) kJ/kg MF coal	29,354	33,010	28,130
Proximate Analysis, wt% MF coal			
Volatile Matter	37.9	36.2	41.2
Fixed Carbon	51.5	59.3	53.6
Ash	10.6	4.5	5.2
Ultimate Analysis, wt% MF coal			
Carbon	70.9	80.5	70.5
Hydrogen	4.9	4.7	4.4
Nitrogen	1.5	1.6	1.3
Sulfur	3.5	1.6	0.5
Oxygen	8.6	7.1	18.1
Ash	10.6	4.5	5.2
Petrographic Analysis, vol%			
Vitrinite		64.0	71.4
Exinite	N/A	17.0	7.3
Inertinite		14.0	18.2
Mineral Matter		5.0	3.1
Ash Analysis, wt%			
Silica, SiO_2	47.8	32.9	39.7
Alumina, Al_2O_3	18.1	27.6	14.6
Titania, TiO_2	0.9	0.8	8.5
Ferric Oxide, Fe_2O_3	17.8	32.2	30.3
Lime, CaO	7.4	2.3	4.5
Magnesia, MgO	0.9	1.0	1.0
Potassium Oxide, K_2O	1.8	1.0	0.9
Sodium Oxide, Na_2O	0.5	1.5	0.2
Phosphorous Pentoxide, P_2O_5	0.01	0.6	0.4
Sulfur Trioxide, SO_3	3.7	-	-
Undetermined	1.1	0.1	-

Notes/Sources:

1. From Burning Star Mine (EPRI Report No. AF-1143-SR, July 1979)
2. From Ruhr area, F-R Germany; also referred to as "standard" German coal by the process developer. (Langhoff et al, paper presented at 29th Canadian Chemical Engineering Conference at Sarnia, Sept. 30, 1979)
3. Coal analysis performed by Birtley Coal and Mineral Testing, and Cascade Coal Petrography Ltd., for SNC Inc., Oct. 1979

Table B-1 Analyses of Different Rank Coals Tested in Advanced Direct Coal Liquefaction and Pyrolysis Processes (cont'd)

Process	Rheinbraun	EDS	
Coal(s)	Rhenish Brown Coal[1]	N. Dakota Indian Head[2]	Texas Big Brown[2]
Moisture (as received), wt%	60.0	N/A	33.6
Gross Calorific Value (GCV) kJ/kg MF coal	23,260	N/A	22,100
Proximate Analysis, wt% MF coal			
Volatile Matter	N/A	44.1	44.4
Fixed Carbon	N/A	45.7	39.2
Ash	3.2	10.2	16.4
Ultimate Analysis, wt% MF coal			
Carbon	65.8	63.8	62.0
Hydrogen	5.0	4.7	4.8
Nitrogen	0.7	19.2	14.5
Sulfur	0.3	0.9	1.1
Oxygen	25.0	1.2	1.2
Ash	3.2	10.2	16.4
Petrographic Analysis, vol%			
Vitrinite	N/A	80.5	65.3
Exinite		2.7	16.3
Inertinite		16.8	18.4
Mineral Matter			
Ash Analysis, wt%			
Silica, SiO_2	N/A	26.3	40.6
Alumina, Al_2O_3		10.7	14.4
Titania, TiO_2		0.5	1.2
Ferric Oxide, Fe_2O_3		14.0	7.8
Lime, CaO		16.0	13.2
Magnesia, MgO		4.9	3.7
Potassium Oxide, K_2O		0.9	0.7
Sodium Oxide, Na_2O		8.0	0.5
Phosphorous Pentoxide, P_2O_5		0.4	0.6
Sulfur Trioxide, SO_3		18.4	16.8
Undetermined		-	0.4

Notes/Sources:

1. J.A. Mikhlin, SNC Inc., visit to Rheinbraun Facilities at Koln, Feb. 19, 1981
2. DOE Report FE-2893-35, p. 18, Sept. 1979

Table B-1 Analyses of Different Rank Coals Tested in Advanced Direct Coal Liquefaction and Pyrolysis Processes (cont'd)

Process	LSE		L-R
Coal(s)	U.K. Bituminous CRC 702[1]	Australian Brown[2]	German Subbituminous
Moisture (as received), wt%	10.0	45.0	4.9
Gross Calorific Value (GCV) kJ/kg MF coal	34,300	23,260	N/A
Proximate Analysis, wt% MF coal			
Volatile Matter	37.7	41.4	38.5
Fixed Carbon	51.2	41.4	56.6
Ash	11.1	17.2	4.9
Ultimate Analysis, wt% MF coal			
Carbon	74.7	62.1	N/A
Hydrogen	4.8	3.6	
Nitrogen	1.7	0.4	
Sulfur	0.6	0.2	
Oxygen	7.1	16.5	
Ash	11.1	17.2	
Petrographic Analysis, vol%			
Vitrinite	N/A	N/A	N/A
Exinite			
Inertinite			
Mineral Matter			
Ash Analysis, wt%			
Silica, SiO_2	N/A	N/A	N/A
Alumina, Al_2O_3			
Titania, TiO_2			
Ferric Oxide, Fe_2O_3			
Lime, CaO			
Magnesia, MgO			
Potassium Oxide, K_2O			
Sodium Oxide, Na_2O			
Phosphorous Pentoxide			
Sulfur Trioxide, SO_3			
Undetermined			

Notes/Sources:

1. NCB Report, "Coal Liquefaction for Transport Fuel Using the Solvent Extraction Technology of U.K. National Coal Board", Feb. 1981
2. W.P. Syrett, NCB, Personal Communication to J.A. Mikhlin, SNC Inc., NCB Ref. DR 3000, DR 496/2, Oct. 1981
3. R.W., Rammler, "The Lurgi-Ruhrgas Process for Carbonization of Coal", paper presented at the EPRI Conference on Coal Pyrolysis, Palo Alto, California, Feb.25-26, 1981.

APPENDIX C

CORRESPONDENCE WITH PROCESS DEVELOPERS

snc

Appendix C-1: Letter from J.L. Gendler, Hydrocarbon Research Inc. - Process Developer of H Coal Process

Hydrocarbon Research Inc. Research & Development Division 134 Franklin Corner Road P.O. Box 6047 Lawrenceville NJ 08648 Tel. 609 896-1300

10-JLG-165-81
October 2, 1981

Dr. Josef A. Mikhlin
Section Head
Coal Research & Development
SNC, Incorporated
1, Complexe Desjardins
P.O. Box 10
Montreal, Canada H5B 1C8

Dear Josef:

I am writing this letter in answer to your request for information on H-Coal® processing of Saskatchewan lignite coal. As I indicated in our telephone conversation, HRI did some bench scale work on a similar lignite (North Dakota) over fifteen years ago, when the program was under the sponsorship of the United States Office of Coal Research (long before the days of the Department of Energy). With the hope that this information will be helpful to your evaluation of our process with respect to the particular feed coal in which you are interested, I am enclosing a copy of that section of our report to the Office of Coal Research which contains the work we did on lignite. To the best of my knowledge, this is the only reported work done by HRI on the subject of lignite liquefaction. We have not run any microautoclave tests on lignite coal.

You asked about HRI's willingness to run microautoclave tests for you, on your specific feed coal. Yes, HRI is willing to run these tests for you, but I must caution you again that the microautoclave is, at best, an imprecise tool by which we can estimate the liquefaction characteristics of a new coal, compared to a similar coal with which we have bench unit or PDU experience. I remind you that our only data on lignite which we could use for comparison is the early bench scale work which I refer to in the paragraph above. (This is the opposite of the situation we faced when testing a subbituminous coal for you last year. We were able to draw the conclusions we did on the basis for comparison with a much more extensive data base.) I do not

Appendix C-1: Letter from J.L. Gendler, Hydrocarbon Research Inc. - Process Developer of H Coal Process (Cont'd).

Dr. Josef A. Mikhlin
October 2, 1981
Page -2-

wish to turn you away from running the microautoclave tests; I merely wish to caution against your using it as any more than a preliminary and non-exact evaluation tool. The financial terms and other conditions for running these tests remain as described in our proposal to you of earlier this year. Although originally quoted for a period of 90 days, we will keep the prices quoted in that proposal in effect through the end of 1981.

In addition to the microautoclave test, if you choose to proceed with such, I would suggest an additional step. We have, for other clients, run microautoclave tests, after which an engineering and economic evaluation was performed. Rather than describe the procedures in this letter, I am enclosing for your perusal a copy of a report recently send to the Government of New South Wales (Australia), in which Australian brown coal was evaluated. In this report, you will find an example of the kind of work we do in our engineering and economic evaluation program. In addition, you may find the data on brown coal useful to your study efforts. The cost of this study, excluding microautoclave tests, is approximately U.S. $25,000.00. The exact price will depend on the number of coal samples tested, the detail into which the report is to go, and so on. Upon your request, I will send you a formal price quotation, along with a schedule and other terms and conditions (by telex if you need it in a hurry).

Of course, the proper way of evaluating the H-Coal® process as applied to Saskatchewan lignite is in bench scale apparatus. If the results of the preliminary study in which you are presently engaged are affirmative, in that H-Coal is selected as one of the more attractive processes for further consideration, I hope that you will continue to work closely with us. A bench scale testing program should be considered a crucial first step before proceeding with detailed engineering and/or economic evaluations. In any event, I wish you success with your study program, and hope that you will not hesitate to call if HRI can be of further service.

Sincerely,

Jeffrey L. Gendler

Jeffrey L. Gendler
HRI R&D Marketing

JLG/jah

Enclosure

Appendix C-1: Excerpt from HRI Report on "H-Coal Evaluation for the Energy Authority of New South Wales", August 1981.

HYDROCARBON RESEARCH, INC.

H-COAL EVALUATION FOR THE ENERGY AUTHORITY OF NEW SOUTH WALES

1.2 - ESTIMATE OF H-COAL YIELDS

As indicated previously, microautoclave testing does not provide definitive yields which can be achieved with Muswellbrook coal in the H-Coal process. In order to estimate H-Coal yields from Muswellbrook coal, it must be compared to a similar coal in the H-Coal data base. The similar coal must be shown to have similar qualities and comparable microautoclave conversions. Once these prerequisites are satisfied, H-Coal yields can be assumed to be the same based on HRI's bench unit or PDU operations. It must be cautioned, however, that even though the two coals may have similar microautoclave test results and qualities, Muswellbrook coal may react differently in the H-Coal process when demonstrated at the bench or PDU scale.

Muswellbrook coal has been shown to react similarly to the client coal as shown in Section 1. The coal qualities for the two coals are summarized below.

Coal Qualities

Coal	Muswellbrook	Client X
Proximate Analysis, Wt% (Dry)		
Ash	22.6	26.1
Volatile Matter	30.6	26.6
Fixed Carbon	46.8	47.3
Ultimate Analysis, Wt% (MAF)		
Carbon	81.9	82.6
Hydrogen	4.8	5.3
Nitrogen	1.9	1.0
Sulfur	0.8	2.0
Oxygen	10.6	9.1
Macerals, Vol %		
Vitrinite	69	-
Exinite	5	-
Inertinite	19	-
Minerals	7	-

The Muswellbrook and client coals have very similar proximate analyses. Both coals have high ash contents, relatively low amounts to volatile matter and similar fixed carbon contents. Comparing the ultimate analysis of the two coals, reveals some differences. The most notable difference is a lower hydrogen content for the Muswellbrook coal. This lower hydrogen content is evident from the maceral composition of the Muswellbrook coal. The high concentration of inertinite, a hydrogen deficient component in coals, causes the low hydrogen content of the sample. It is interesting to note that if the maceral composition is converted to weight percent, on a mineral free basis, the inertinite is roughly 30 percent of the mineral free coal. The ultimate conversion in the microautoclave test, on an MAF basis was 73 percent. This indicates that the vitrinite portion of the Muswellbrook coal is highly reactive, and highly desirable for coal liquefaction.

The Muswellbrook coal, as received by HRI, would probably not be economic for commercial coal liquefaction. This is due to its high ash content (22.6 Wt percent on dry coal) and high inertinite content (19 Vol percent, roughly 30 Wt%). The high solids content of the H-Coal reactor product would require large amounts of coal liquids to yield a pumpable bottoms stream (about 50 percent solids). The H-Coal evaluation for Muswellbrook coal is based on coal cleaned to 10 percent ash. The coal cleaning will also reduce the inertinite content of the coal. The total amount of inertinite plus ash should be on the order of 10-20 weight percent for economic operations.

The H-Coal yields for Muswellbrook coal are based on HRI's PDU operations on the client coal, cleaned to a low ash level (about 15 weight percent). The yields were assumed to be similar on an MAF basis. The hydrogen consumption was increased to reflect the lower hydrogen content of Muswellbrook coal and to maintain the same product qualities. Adjustments were made to the yields of NH_3, H_2S H_2O and CO_x to reflect differences in the elemental compositions. The estimated H-Coal yields for Muswellbrook coal are:

Estimated H-Coal Yields

Yield	Wt% on MAF Coal
H_2	5.72
CO	0.11
CO_2	0.30
H_2S	0.55
NH_3	1.04
H_2O	7.63
C_1/C_2	7.89
C_3	3.57
C_4	3.99
C_5/180°F	3.87
180/350°F	8.12
350/700°F	21.15
700/800°F	1.86
800°F + Unconverted Coal	45.86
Total	100.00

HYDROCARBON RESEARCH, INC.

H-COAL EVALUATION FOR THE ENERGY AUTHORITY OF NEW SOUTH WALES

Although Muswellbrook coal has been shown to be similar to a coal in the H-Coal data base, it may react differently in the H-Coal process when demonstrated at the bench or PDU scale. In addition, no attempt has been made to obtain optimum operating conditions, and thereby an optimum yield slate for Muswellbrook coal. Further evaluation of Muswellbrook coal, especially a cleaned Muswellbrook coal, in HRI's bench units or PDU, as well as research leading to an improved H-Coal process, may significantly improve H-Coal yields

Appendix C-2: Telex from Veba Oel Ag - A Process Developer of RAG/Veba Process

TELEX-NR. 260/553/HS-KUE

PROF. DR. J. A. MIKHLIN
SNC

1981 NOV 20 AM 10: 11

HYDROGENATION OF WILLOW BUNCH OR ESTEVAN COAL

DEAR DR. MIKHLIN,

AFTER YOUR LAST VISIT WE HAD SOME FURTHER DISCUSSIONS WITH OUR DR. GRAESER AND DR. NIEMANN REGARDING THE POSSIBILITIES OF HYDROGENATION OF WILLOW BUNCH OR ESTEVAN COALS. FURTHERMORE WE DID SOME CALCULATIONS. FOLLOWING ARE THE RESULTS:

BOTH COALS HAVE TWO DISADVANTAGES FOR HYDROGENATION COMPARED TO GERMAN STANDARD COAL.

- THE HIGH ASH CONTENT
- THE HIGH OXYGEN CONTENT

THE HIGH OXYGEN CONTENT RESULTS IN PRODUCING WATER DURING HYDROGENATION WHICH NEEDS EXPENSIVE HYDROGEN. BUT THIS POINT CAN NOT BE INFLUENCED.

THE HIGH ASH CONTENT RESULTS IN GASIFICATION OF A HIGH AMOUNT OF HYDROGENATION PRODUCTS. THEREFORE THE ASH CONTENT SHOULD BE REDUCED WHICH IS POSSIBLE IN MOST CASES BY WASHING THE RAW COAL USING SPECIAL METHODS.

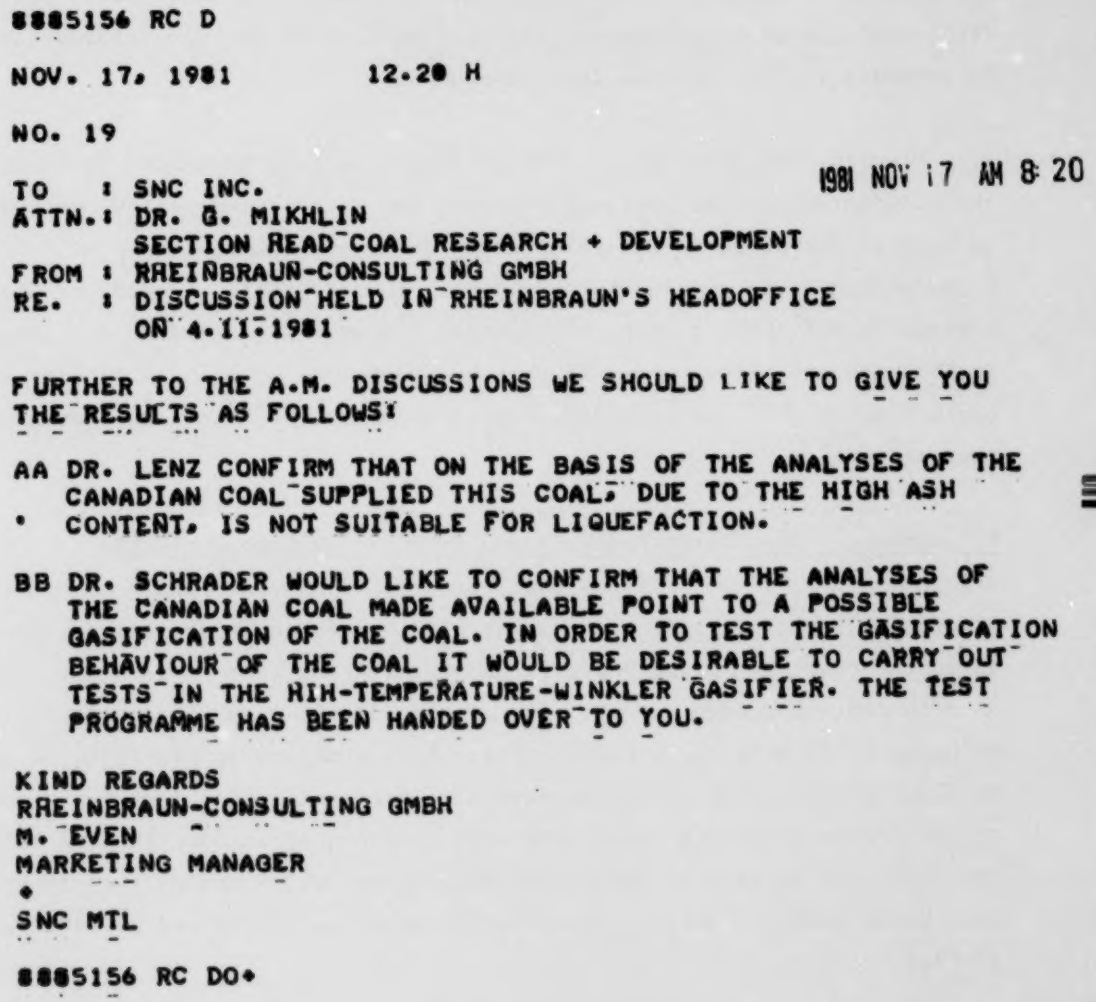
THE YIELDS AS WE HAVE CALCULATED TOGETHER DURING YOUR VISIT REMAIN IN THE SAME RANGE.

DUE TO THESE FACTS WE WOULD RECOMMEND TO GASIFY THE COAL IF REDUCTION OF ASH CONTENT IS NOT POSSIBLE. ON THE OTHER HAND YOU SHOULD TRY TO FIND OUT POSSIBILITIES TO WASH THE COAL. IF REDUCTION OF ASH CONTENT TO LESS THAN 10 PERCENT IS POSSIBLE WE WOULD RECOMMEND HYDROGENATION TESTS IN OUR AUTOCLAVE UNIT.

BEST REGARDS
VEBA OEL ENTWICKLUNGS-GMBH
553 - COAL CONVERSION
SCHMEDESHAGEN

+++
824639 VOE D+
SNC MTL

Appendix C-3: Telex from M. Even, Rheinbraun - Consulting GMBH - Process Developer of Rheinbraun Coal Liquefaction Process and High-Temperature Winkler Gasifier.



8885156 RC D

NOV. 17, 1981 12.20 H

NO. 19

1981 NOV 17 AM 8:20

TO : SNC INC.
ATTN.: DR. G. MIKHLIN
SECTION HEAD COAL RESEARCH + DEVELOPMENT
FROM : RHEINBRAUN-CONSULTING GMBH
RE. : DISCUSSION HELD IN RHEINBRAUN'S HEADOFFICE
ON 4.11.1981

FURTHER TO THE A.M. DISCUSSIONS WE SHOULD LIKE TO GIVE YOU THE RESULTS AS FOLLOWS:

AA DR. LENZ CONFIRM THAT ON THE BASIS OF THE ANALYSES OF THE CANADIAN COAL SUPPLIED THIS COAL, DUE TO THE HIGH ASH CONTENT, IS NOT SUITABLE FOR LIQUEFACTION.

BB DR. SCHRADER WOULD LIKE TO CONFIRM THAT THE ANALYSES OF THE CANADIAN COAL MADE AVAILABLE POINT TO A POSSIBLE GASIFICATION OF THE COAL. IN ORDER TO TEST THE GASIFICATION BEHAVIOUR OF THE COAL IT WOULD BE DESIRABLE TO CARRY OUT TESTS IN THE HIH-TEMPERATURE-WINKLER GASIFIER. THE TEST PROGRAMME HAS BEEN HANDED OVER TO YOU.

KIND REGARDS
RHEINBRAUN-CONSULTING GMBH
M. EVEN
MARKETING MANAGER
+
SNC MTL

8885156 RC DO+

Appendix C-4 Minutes of the Meeting Between SNC and Exxon Research and Engineering Company held by Telephone on October 26, 1981

Agenda

Suitability of Saskatchewan lignites for processing in Exxon Donor Solvent (EDS) coal liquefaction process and availability of process information to be supplied to SNC for technical assessment.

J.A. Mikhlin, SNC (Montreal) Some time ago, we sent by telex to Exxon, the proximate, ultimate and ash analyses for Estevan and Willow Bunch lignites as well as discussed in the telex the program on Saskatchewan Coal Liquefaction Study awarded to SNC by SASKOIL and CANMET. Would you please comment on suitability of the designated Saskatchewan lignites for liquefaction in EDS process. Also, when and what documentation or process package SNC could receive from you in order to make objective technical assessment of your process.

H.F. Corning, Jr., Exxon Research and Engineering Company (ER&E) (Florham Park, N.J.) As per your request, we have already sent you the information available on processing lignites in our PDU unit as well as other updated information. Exxon Research and Engineering Company is right now working on a conceptual design up-date of a commercial EDS plant for processing of bituminous Illinois No. 6 coal. After this, sometime in mid-1982, we intend to develop conceptual design up-date of a commercial EDS plant for processing of subbituminous Wyodak coal. The work on a conceptual design of a commercial EDS plant for processing Texas lignite, the one which would be of interest for your Saskatchewan lignites, should start sometime at the end of 1982 or beginning of 1983.

As you know, in accordance to the experimental program being carried out at 250 t/d EDS pilot plant in Baytown, Texax, the Texas lignite should be tested in 1982, after the testing programs with Illinois No. 6 and Wyodak coals are completed. This experimental program will include the testing of the process in the bottoms-recycle mode which should have very favourable effect on product yields specifically for processing high-ash lignites such as Texas or, as in your case, for Saskatchewan lignites.

We intend to include the Baytown pilot plant data in our process design update work.

J.A. Mikhlin, SNC (Montreal) From your point of view, why would bottoms recycle mode of EDS process be favourable especially for liquefaction of high-ash coals such as Texas or Saskatchewan lignites?

H.F. Corning, Jr. (ER&E) Bottoms-recycle mode would lower the viscosity of distillation bottoms and thus reduce the problems of processing high ash content coal. Also, bottoms recycle should increase light liquid product yield (naphtha, mid-distillate) on account of heavy oil produced.

J.A. Mikhlin, SNC (Montreal) What is your estimate of the time it will take ER&E to analyse the Baytown Pilot Plant data and information before it may be used in your work on design up-date cases?

H.F. Corning, Jr. (ER&E) It takes usually about six months at least after the completion of a pilot plant experimental program or part of it related to a certain coal testing before the results of the program may be used for process design such as conceptual design up-date of a process plant.

J.A. Mikhlin, SNC (Montreal) Thank you very much for the information. I will continue to be in touch with you as our work on Saskatchewan Coal Liquefaction Study proceeds.

Appendix C-5: Telex from W.B. Crouch, Texaco Development Corporation - Process Developer of Texaco Coal Gasification Process.

1982 JAN 12 PM 1:16

SNC INC
MONTREAL (CANADA)
TELEX: 055-60042
ATTN: DR. J. A. MIKHLIN
JANUARY 11, 1982
RE: SNC STUDY, SASK LIGNITE

AA. PRELIMINARY ESTIMATE OF OPERATION FOR THE TEXACO COAL GASIFICATION PROCESS, BASED UPON PARTIAL WILLOWBUNCH LIGNITE ANALYTICAL TEST DATA YOU PROVIDED DURING YOUR OCTOBER 22, 1981 VISIT TO TDC OFFICES PLUS ASSUMPTIONS FROM TESTS ON OTHER SIMILAR COALS, (AS SPECIFIED ITEM 2 MY DECEMBER 22 TELEX TO YOU), FOLLOWS. WISH TO REPEAT OUR CAUTION THAT THIS IS PRELIMINARY PROPOSAL BASED UPON INCOMPLETE INFO FOR YOUR PRELIMINARY SCOPING AND WILL BE MODIFIED AFTER TEST RESULTS YOUR LIGNITE AVAILABLE.

TEXACO DEVELOPMENT CORPORATION
COAL GASIFICATION PROCESS
ESTIMATE OF OPERATION
JAN 11, 1982

FOR - SNC STUDY
LOCATION - SASKATCHEWAN, CANADA
CHARGE STOCK - WLLOWBUNCH LIGNITE
ULTIMATE PRODUCT - METHANOL OR FISCHER-TROPSCH FEED

	NORMAL OPERATING CONDITION
CHARGE TO GENERATOR	
HYDROCARBON (FRESH, DRY), SHORT TONS PER OPERATING DAY POUNDS	3116.2
POUNDS PER HOUR	259683.
ULTIMATE ANALYSIS, WEIGHT PERCENT, MOISTURE FREE	
CARBON	52.10
HYDROGEN	4.00
NITROGEN	0.50
SULFUR	0.90
OXYGEN	21.50
ASH	21.00
HIGHER HEATING VALUE, BTU PER POUND	8372.
PUR OXYGEN, SHORT TONS PER OPERATING DAY	2497.9
PURITY OF OXIDANT STREAM, MOL PERCENT OXYGEN	99.5
WATER, POUNDS PER HOUR	281323.

Appendix C-5: Telex from W.B. Crouch, Texaco Development Corporation - Process Developer of Texaco Coal Gasification Process (Cont'd)

PRODUCT COMPOSITION, MOL PERCENT

CARBON MONOXIDE	15.32
HYDROGEN	18.94
CARBON DIOXIDE	19.71
WATER	45.48
METHANE	0.08
ARGON	0.10
NITROGEN	0.14
HYDROGEN SULFIDE	0.22
CARBONYL SULFIDE	0.01 (.009)

HYDROGEN PLUS CARBON MONOXIDE,

MILLION STANDARD CUBIC FEET -PE OPERATING DAY	100.000
SLAG, LB./HR.	38306.
CARBON CONTENT, WEIGHT PERCENT	0.
CHAR DISCHARGE, LB./HR.	15545.
CARBON CONTENT, WEIGHT PERCENT	0.

SCF - STANDARD CUBIC FEET MEASURED AT 60F AND 14.696 POUNDS PER SQUARE INCH ABSOLUTE.

WBC: 8607A

BB. SHIPMENT OF SAMPLES TO TEXACO BEACON RESEARCH LABORATORY
FROM OUTSIDE UNITED STATES:
AIR SHIPMENT TO BEACON RESEARCH LABORATORIES
TEXACO INC.
OLD GLENHAM ROAD
GLENHAM, NEW YORK 12527
ATTN: DR. R. B. TIERNEY
FROM OUTSIDE U. S. SHOULD BE ADDRESSED TO
QUOTE TEXACO INTERNATIONAL TRADER INC.
C/O DANIEL F. YOUNG INC.
176-20 147 AVENUE
JAMAICA, NEW YORK 11434
NOTIFY SAME (AND TEXACO BEACON)
ON ARRIVAL UNQUOTE
ORIGINAL AIR BILL AND COMMERCIAL INVOICE MUST ACCOMPANY SHIPMENT. SEND COPY OF ALL DOCUMENTS TO:

MR. J. C. WALLIN
TEXACO INTERNATIONAL TRAFFIC DEPT.
P. O. BOX 52332
HOUSTON, TEXAS 77052

BEST REGARDS
W. B. CROUCH
TEXACO DEVELOPMENT

DTX HAA 01/12 11:59 3218
1351 TXO 01/12 12:02

Appendix C-5: Telex from W.B. Crouch, Texaco Development Corporation - Process Developer of Texaco Coal Gasification Process

1982 FEB -8 PM 4:51

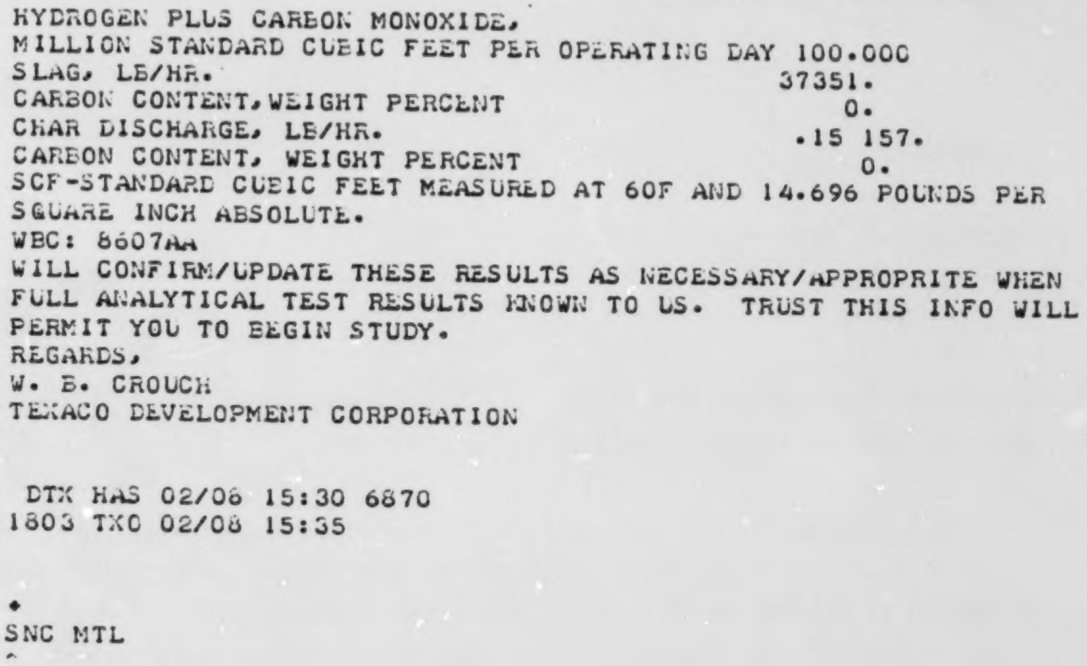
SNC INC
MONTREAL (CANADA)
TELEX: 055-60042
ATTN: DR. J. A. MIKHLIN
FEBRUARY 8, 1982
RE: SNC STUDY, SASK LIGNITE

PRELIMINARY ESTIMATE OF OPERATION FOR THE TEXACO COAL GASIFICATION PROCESS, (TCGP) BASED UPON PARTIAL WILLOWBUNCH LIGNITE ANALYTICAL TEST DATA YOU PROVIDED DURING YOUR OCTOBER 22, 1981 VISIT TO TDC OFFICES PLUS PRELIMINARY (SLURRY CONC ONLY) TESTS OR LIGNITE SAMPLES AFTER VACUUM-DRYING (WHICH WE ASSUME SIMULATED COMMERCIALLY-AVAILABLE LIGNITE-DRYING) SLURRY CONC OF 49.5 PCT DRY SOLIDS ATTAINED. OUR TCGP ESTIMATE OF OPERATION BASED ON THIS RESULT, FOLLOWS:

TEXACO DEVELOPMENT CORPORATION
COAL GASIFICATION PROCESS
ESTIMATE OF OPERATION
FEBRUARY 8, 1982
FOR - SNC
LOCATION- SASKATCHEWAN, CANADA
CHARGE STOCK - WILLOWBUNCH LIGNITE
ULTIMATE PRODUCT - METHANOL OR FISCHER-TROPSCH FEED

	NORMAL OPERATING CONDITION
CHARGE TO GENERATOR	
HYDROCARBON (FRESH, DRY), SHORT TONS PER	
OPERATING DAY PC	3038.5
POUNDS PER HOUR	.253208
ULTIMATE ANALYSIS, WT PCT, MOISTURE FREE	
CARBON	52.10
HYDROGEN	4.00
NITROGEN	0.50
SULFUR	0 .90
OXYGEN	21.50
ASH	21.00
HIGHER HEATING VALUE, BTU PER POUND	8372.
PURE OXYGEN, SHORT TONS PER OPERATING DAY	2363.1
PURITY OF OXIDANT STREAM, MOL PCT OXYGEN	99.5
WATER, POUNDS PER HOUR	281323.
PRODUCT COMPOSITION, MOL PERCENT	
CARBON MONOXIDE	16.47
HYDROGEN	19.69
CARBON DIOXIDE	19.56
WATER	43 .69
METHANE	0 .0
ARGON	0.10
NITROGEN	0.15
HYDROGEN SULFIDE	0.22
CARBONYL SULFIDE	0.01(.009)

Appendix C-5: Telex from W.B. Crouch, Texaco Development Corporation - Process Developer of Texaco Coal Gasification Process (Cont'd)



```
HYDROGEN PLUS CARBON MONOXIDE,
MILLION STANDARD CUBIC FEET PER OPERATING DAY 100.000
SLAG, LB/HR.                                   37351.
CARBON CONTENT,WEIGHT PERCENT                      0.
CHAR DISCHARGE, LB/HR.                          .15 157.
CARBON CONTENT, WEIGHT PERCENT                     0.
SCF-STANDARD CUBIC FEET MEASURED AT 60F AND 14.696 POUNDS PER
SQUARE INCH ABSOLUTE.
WBC: 8607AA
WILL CONFIRM/UPDATE THESE RESULTS AS NECESSARY/APPROPRITE WHEN
FULL ANALYTICAL TEST RESULTS KNOWN TO US.  TRUST THIS INFO WILL
PERMIT YOU TO BEGIN STUDY.
REGARDS,
W. B. CROUCH
TEXACO DEVELOPMENT CORPORATION


 DTX HAS 02/06 15:30 6870
1803 TXC 02/08 15:35


◆
SNC MTL
```

Appendix C-6 Minutes of the Meeting Held at Lurgi Kohle and Mineraloltechnik Offices in Frankfurt (Main) on November 5, 1981

Agenda

Discussion on suitability of commercial Lurgi dry bottom gasifiers as a first step in a indirect coal liquefaction complex using Estevan or Willow Bunch lignites as a feedstock.

Present

Dr.-Ing. P.K. Herbert	Lurgi, F.R.G.
Dr.-Ing. R. Rammler	Lurgi, F.R.G.
Dr. J. Mikhlin	SNC, Canada

Discussion was based on the proximate, ultimate and ash analyses of Estevan and Willow Bunch lignite submitted to Lurgi by SNC.

The Lurgi gasifier is a first generation pressurized, fixed-bed, dry-bottom gasifier which has been commercially proven on high-ash-content South African lignites (27 wt% ash on MF basis). Also, some American lignites were successfully tested in Lurgi gasifier. Therefore, normal operation is assumed by Lurgi if Saskatchewan lignites will be gasified in this gasifier.

If the Lurgi gasifier is selected for syngas production, substantial amount of methane in the raw gas will be produced and some liquid hydrocarbon by-products may be expected. In order to maximize the syngas production, methane must be separated and reformed to syngas. An alternative is to use the methane separated from raw gas as a fuel.

Large amounts of fines expected in the Saskatchewan lignite (around 25%) would require fines separation. Coal fines may be burned in the utilities or briquetted to a desired size before feeding into the gasifier.

Saskatchewan lignites may be gasified in the Lurgi dry-bottom gasifier after the extra fines are separated. However, to define the actual raw gas composition, coal devolatilization under pressure, behaviour of ash in the gasifier, oxygen and steam consumption, etc., a coal gasification test using representative samples of Estevan and Willow Bunch lignites will be required.

Lurgi dry-bottom gasifiers are robust and stable in operation. Little CO_2 shifting, if any, is required before feeding the cleaned raw gas into the subsequent synthesis reactor.

Appendix C-7: Telex from B.H. Thompson, British Gas Corporation - A Process Developer of British Gas/Lurgi Slagging Gasifier.

F.A.O: DR J MIKHLEN
SECTION HEAD COAL RESEARCH AND DEVELOPMENT
SNG INC.

1982 JAN 11 PM 1:10

FROM: DR B H THOMPSON
ICS
BRITISH GAS

BRITISH GAS/LURGI SLAGGING GASIFIER

WE HAVE NOW HAD THE OPPORTUNITY TO CONSIDER THE SUITABILITY OF THE SASKATCHEWAN COALS FOR GASIFICATION IN THE SLAGGING GASIFIER.

THE DATA YOU PROVIDED GAVE US SOME PROPERTIES FOR THE ESTEVAN AND WILLOW BUNCH COALS, BUT DETAILED DATA SUCH AS ASH ANALYSIS WAS ONLY GIVEN FOR THE FORMER COAL. WE IDENTIFY THE ESTEVAN NONE COAL AS A TYPE A LIGNITE CONTAINING SIGNIFICANT AMOUNTS OF MOISTURE (35□ A.R.) AND ASH (13.65□ A.R.). THE ASH IS NOT VERY REFRACTORY AND ANY FLUX ADDITION NECESSARY TO MODIFY THE ASH CHARACTERISTICS WOULD BE SMALL.

WE ARE OF THE OPINION THAT THESE LIGNITES SHOULD BE SUITABLE FOR GASIFICATION IN THE SLAGGING GASIFIER, BUT IN ORDER TO OBTAIN A SENSIBLE OUTLET GASIFIER TEMPERATURE, IT WOULD PROBABLY BE NECESSARY TO DRY THE LIGNITE TO SOME EXTENT.

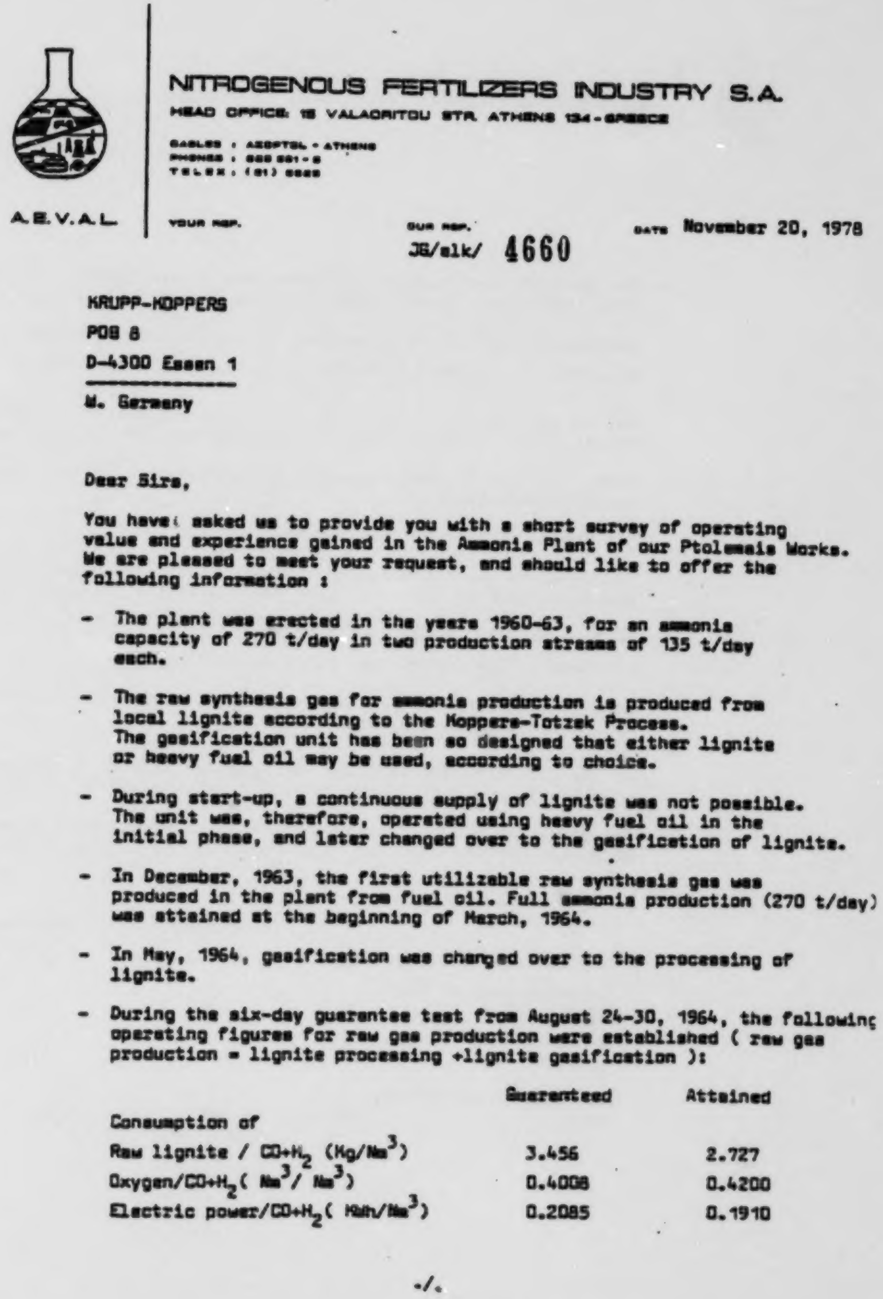
ALSO, DUE TO LACK OF OPERATIONAL EXPERIENCE ON GASIFYING LIGNITES ANY DATA WE PRODUCED WOULD ONLY BE SUITABLE FOR PRELIMINARY DESIGN EVALUATION AND PLANT OPERATION AT WESTFIELD WOULD BE NECESSARY TO OBTAIN GOOD DESIGN DATA WITH A SUFFICIENT SUPPLY OF THE MATERIAL DELIVERED TO OPERATE THE PLANT FOR AN APPROPRIATE PERIOD OF TIME. WE ARE PREPARED TO CONSIDER THIS UNDER COMMERCIAL TERMS.

PLEASE ACCEPT MY APOLOGIES FOR THE DELAY IN RESPONSE TO YOUR ENQUIRY. I THOUGHT I HAD RESPONDED EARLIER BUT ON CHECKING MY RECORDS, I FOUND THAT I HAD NOT.

BEST REGARDS,
DR B H THOMPSON
ICS
BRITISH GAS.

265812 ICS G
◆
SNC MTL

Appendix C-8: Copy of the letter from J. Gerakis of Nitrogenous Fertilizers Industry S.A. received from Krupp-Koppers GMbH - Process Developer of Koppers-Totzek Gasifier



NITROGENOUS FERTILIZERS INDUSTRY S.A.
HEAD OFFICE: 15 VALAORITOU STR. ATHENS 134 - GREECE
CABLES : AZOPTOL - ATHENS
PHONES : [illegible]
TELEX : (21) [illegible]
A.E.V.A.L.

YOUR REF.

OUR REF.
JG/elk/ 4660

DATE November 20, 1978

KRUPP-KOPPERS
POB 8
D-4300 Essen 1
W. Germany

Dear Sirs,

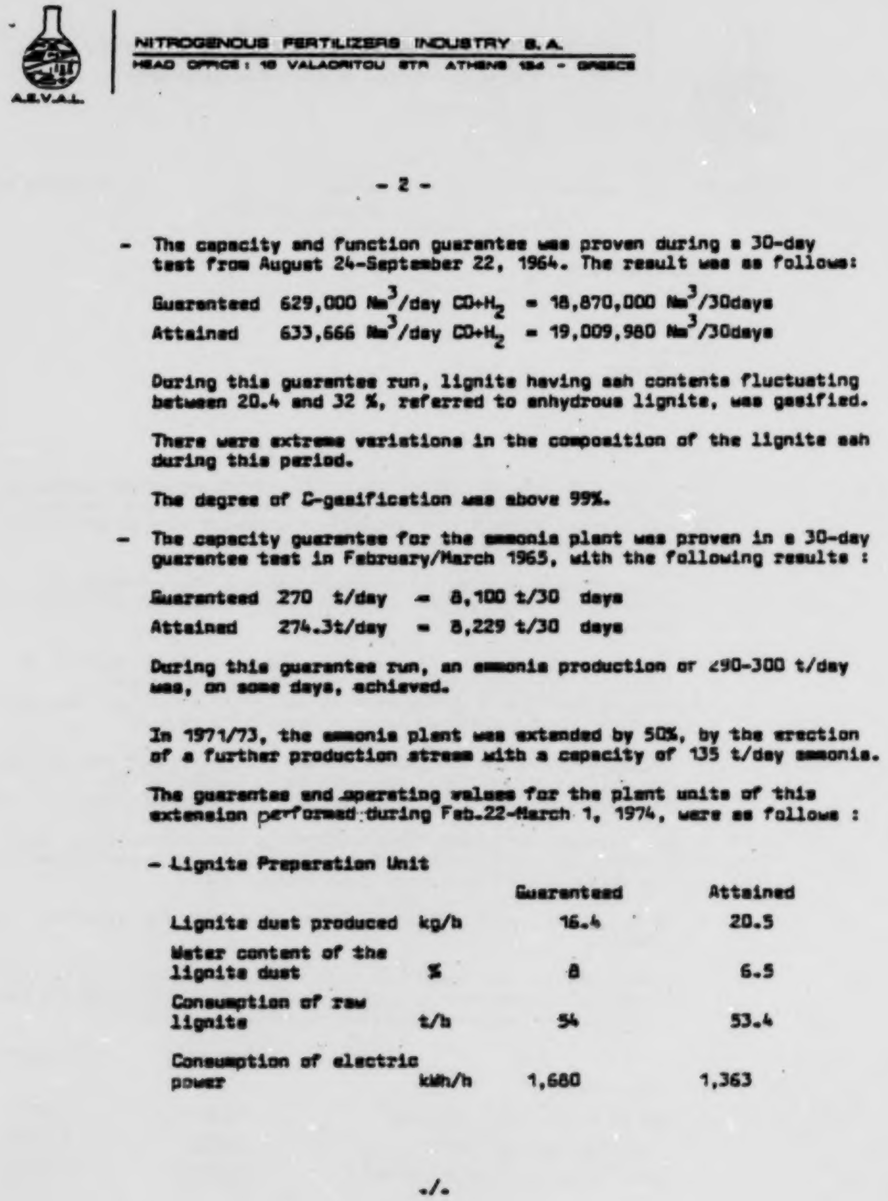
You have asked us to provide you with a short survey of operating value and experience gained in the Ammonia Plant of our Ptolemais Works. We are pleased to meet your request, and should like to offer the following information :

- The plant was erected in the years 1960-63, for an ammonia capacity of 270 t/day in two production streams of 135 t/day each.
- The raw synthesis gas for ammonia production is produced from local lignite according to the Koppers-Totzek Process. The gasification unit has been so designed that either lignite or heavy fuel oil may be used, according to choice.
- During start-up, a continuous supply of lignite was not possible. The unit was, therefore, operated using heavy fuel oil in the initial phase, and later changed over to the gasification of lignite.
- In December, 1963, the first utilizable raw synthesis gas was produced in the plant from fuel oil. Full ammonia production (270 t/day) was attained at the beginning of March, 1964.
- In May, 1964, gasification was changed over to the processing of lignite.
- During the six-day guarantee test from August 24-30, 1964, the following operating figures for raw gas production were established (raw gas production = lignite processing +lignite gasification):

	Guaranteed	Attained
Consumption of		
Raw lignite / $CO+H_2$ (Kg/Nm^3)	3.456	2.727
Oxygen/$CO+H_2$(Nm^3/ Nm^3)	0.4008	0.4200
Electric power/$CO+H_2$(KWh/Nm^3)	0.2085	0.1910

./.

Appendix C-8: Copy of the letter from J. Gerakis of Netrogenous Fertilizers Industry S.A. received from Krupp-Koppers GMbH - Process Developer of Koppers-Totzek Gasifier (Cont'd)



A.E.V.A.L.

NITROGENOUS FERTILIZERS INDUSTRY S.A.
HEAD OFFICE: 10 VALAORITOU STR ATHENS 134 - GREECE

- The capacity and function guarantee was proven during a 30-day test from August 24-September 22, 1964. The result was as follows:

Guaranteed 629,000 Nm^3/day $CO+H_2$ = 18,870,000 Nm^3/30days
Attained 633,666 Nm^3/day $CO+H_2$ = 19,009,980 Nm^3/30days

During this guarantee run, lignite having ash contents fluctuating between 20.4 and 32 %, referred to anhydrous lignite, was gasified.

There were extreme variations in the composition of the lignite ash during this period.

The degree of C-gasification was above 99%.

- The capacity guarantee for the ammonia plant was proven in a 30-day guarantee test in February/March 1965, with the following results :

Guaranteed 270 t/day = 8,100 t/30 days
Attained 274.3t/day = 8,229 t/30 days

During this guarantee run, an ammonia production or 290-300 t/day was, on some days, achieved.

In 1971/73, the ammonia plant was extended by 50%, by the erection of a further production stream with a capacity of 135 t/day ammonia.

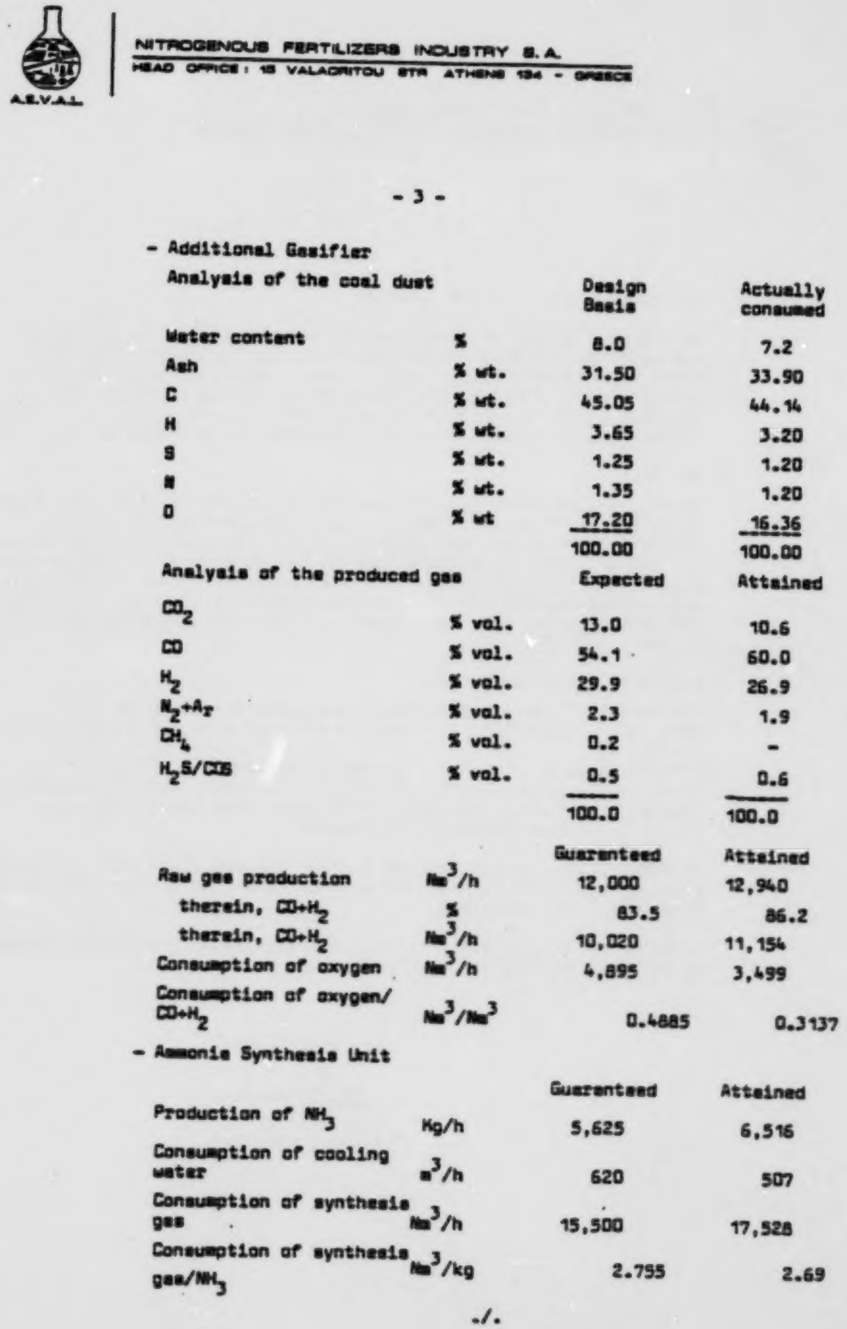
The guarantee and operating values for the plant units of this extension performed during Feb.22-March 1, 1974, were as follows :

- Lignite Preparation Unit

		Guaranteed	Attained
Lignite dust produced	kg/h	16.4	20.5
Water content of the lignite dust	%	8	6.5
Consumption of raw lignite	t/h	54	53.4
Consumption of electric power	kWh/h	1,680	1,363

./.

Appendix C-8: Copy of the letter from J. Gerakis of Nitorgenous Fertilizers Industry S.A. received from Krupp-Koppers GMbH - Process Developer of Koppers-Totzek Gasifier (Cont'd)



A.S.V.A.L.
NITROGENOUS FERTILIZERS INDUSTRY S.A.
HEAD OFFICE: 15 VALAORITOU STR ATHENS 134 - GREECE

- Additional Gasifier

Analysis of the coal dust		Design Basis	Actually consumed
Water content	%	8.0	7.2
Ash	% wt.	31.50	33.90
C	% wt.	45.05	44.14
H	% wt.	3.65	3.20
S	% wt.	1.25	1.20
N	% wt.	1.35	1.20
O	% wt	17.20	16.36
		100.00	100.00

Analysis of the produced gas		Expected	Attained
CO_2	% vol.	13.0	10.6
CO	% vol.	54.1	60.0
H_2	% vol.	29.9	26.9
N_2+A_r	% vol.	2.3	1.9
CH_4	% vol.	0.2	-
H_2S/COS	% vol.	0.5	0.6
		100.0	100.0

		Guaranteed	Attained
Raw gas production	Nm^3/h	12,000	12,940
therein, $CO+H_2$	%	83.5	86.2
therein, $CO+H_2$	Nm^3/h	10,020	11,154
Consumption of oxygen	Nm^3/h	4,895	3,499
Consumption of oxygen/ $CO+H_2$	Nm^3/Nm^3	0.4885	0.3137

- Ammonia Synthesis Unit

		Guaranteed	Attained
Production of NH_3	Kg/h	5,625	6,516
Consumption of cooling water	m^3/h	620	507
Consumption of synthesis gas	Nm^3/h	15,500	17,528
Consumption of synthesis gas/NH_3	Nm^3/kg	2.755	2.69

./.

Appendix C-8: Copy of the letter from J. Gerakis of Nitrogenous Fertilizers Industry S.A. received from Krupp-Koppers GMbH - Process Developer of Koppers-Totzek Gasifier (Cont'd)

A.E.V.A.L.
NITROGENOUS FERTILIZERS INDUSTRY S.A.
HEAD OFFICE: 18 VALAORITOU STR ATHENS 134 - GREECE

- **Refractory Lining of the Gasifier**

  In the past period of operation, we had practically no damage at the refractory lining of the gasifiers. During regular inspections, small repairs were carried out, whereby defective spots, caused by erosion, were replaced using ramming mass. These repairs could be executed within a period of a few hours. In other words, the refractory lining has not needed to be renewed since commencement of start-up.

- **Ash Split-up**

  Depending on the ash content and mode of operation of the gasifiers, about 40-60% of the coal ash is discharged beneath the gasifier in the form of slag. The remaining 60-40%, entrained as fly dust with the gas produced, is carried through the waste heat boiler into the cooling washer, and is, there, washed out. The carbon content in the slag discharged beneath the gasifiers, amounts to about 0.2-0.25%. The carbon in the fly dust, discharged with the gas from the gasifier, is about 2-2.10%

- **Degree of Carbon Conversion.**

  The degree of carbon conversion during the operating period amounted to about 98%. At times, even a value of above 99% was attained.

- Interruptions to operation, caused by breakdown of the Gasification Unit, were extremely rare, in spite of operation with fluctuating lignite qualities. We estimate that these amounted to about 2% of the total operational disturbances.

In summarizing, we can confirm that we are satisfied with the operation and results of the Plant. The planned annual production figures have, in general, been attained.

We agree to this letter's being shown to serious prospective customers or to their being given a copy of this.

Very truly yours,

J. Gerakis
Managing Director.

Appendix C-9: Letter from H.B. Beisel, Synthetic Fuel Process, Westinghouse Canada Inc.

Westinghouse Canada Inc.

Energy Resource Centre, Alberta
Suite 807
The Home Oil Tower
Toronto Dominion Square
Calgary Alberta
T2P 2Z2

SFD-GJK-82-035
February 18, 1982

Dr. Josef A. Mikhlin
The SNC Group
1 Complexe Desjardins
P. O. Box 10
Desjardins Postal Station
Montreal, Canada H5B 1C8

Dear Dr. Mikhlin:

In response to your recent request to the Westinghouse Synthetic Fuels Division, we are pleased to send you a proposal for a conceptual design study of the Westinghouse coal gasification process. The information to be provided in this study is specific to the proposed Saskatchewan Oil and Gas Corporation project in Western Canada to produce coal liquids from lignite.

Prior to undertaking the proposed conceptual design study, it is necessary to perform bench scale analyses on the lignite feedstock in order to assess its applicability to the Westinghouse process. These test results will provide a basis for the heat and material balance calculations in the conceptual design study.

Enclosure 1 is the Proposed Scope of Work for the conceptual design study, describing the information to be provided by Westinghouse. The scope for the coal characterization tests mentioned above is provided in Enclosure 2. Enclosure 3 is the Westinghouse Synthetic Fuels Division's Terms and Conditions for Engineering Services. These terms and conditions provide the basis under which Westinghouse will perform this work, and should be incorporated by reference into your purchase order. This purchase order should be sent to the Westinghouse Energy Resource Centre in Alberta who will have the overall responsibility for administering this order. Also, a confidentiality agreement between Westinghouse and the SNC Group is provided in Enclosure 4. A similar confidentiality agreement will also be required between Westinghouse and Saskatchewan Oil and Gas Corporation (Saskoil) before information can be transmitted by SNC to Saskoil. The price for the proposed study, including the coal characterization tests, is U.S. $44,000. This proposal is valid for ninety (90) days.

Westinghouse will initiate work, beginning with the coal characterization tests, after Westinghouse's receipt and acceptance of a purchase order from SNC, the execution of the enclosed confidentiality agreement, and receipt of the coal sample. The conceptual design study information will be delivered to SNC twelve (12) weeks after the initiation of work. Preliminary heat and material balances will be delivered six (6) weeks after the initiation of work.

We look forward to working together with you on this study. If you have any questions, please do not hesitate to contact me, Mr. Keane, or Mr. Dominicis.

Very truly yours,

Harold M. Beisel, Manager
Synthetic Fuels Processes

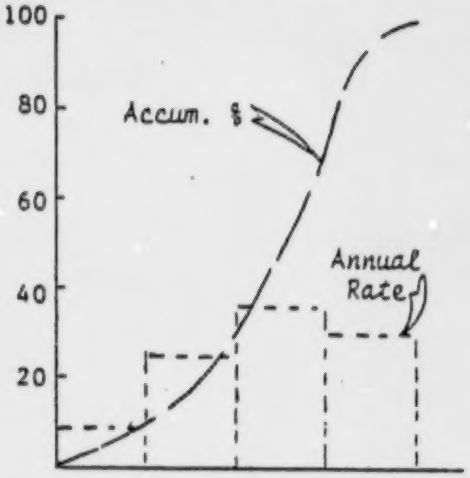
Enclosures

cc: D. P. Dominicis
G. J. Keane

APPENDIX D

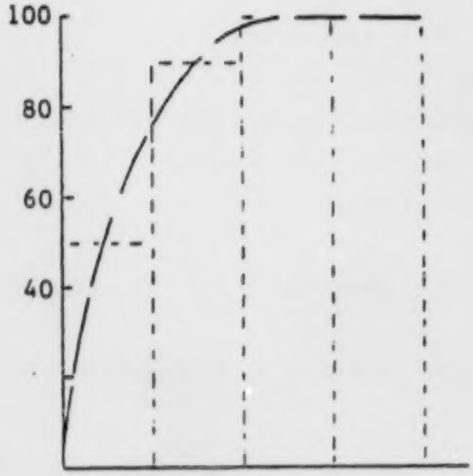
ECONOMIC AND FINANCIAL ASSUMPTIONS AND CRITERIA

Appendix D-1 Capital Investment Schedule (Spending Schedule)



Period*	-4	-3	-2	-1
% Year	9	25	36	30
Accum. %	9	34	70	100

Appendix D-2 Production Level (Percentage of Annual Production Rate)



Period*	1	2	3	4
Production level,%	50	90	100	100

---

* With reference to plant start-up date.

Source: B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979

Appendix D-3: Canada/U.S. Material Price Differences

	U.S.A.	Canada	Variance, %
Concrete Mix, U.S. $/cwt	47.90	50.12	4.6
Standard Structures, U.S. $/cwt			
Wide Flanges	26.20	30.95	18.13
I-Beams	27.70	33.95	22.6
Piping, U.S. $/ft			
Iron	4.51	6.44	42.8
Copper	0.81	0.55	32.1
Cables/Wires			10

Assuming the material distribution of:

30% concrete
15% steel (50% wide flanges, 50% I-beams)
45% piping (80% iron, 20% copper)
10% cables and wires

the "Material Difference Factor" of 1.1592 is obtained.

Therefore, "Material plus Currency Differential Factor" is 1.1592 x 1.19
= 1.379

Source: Engineering News Record, Jan. 7, 1982; Jan. 14, 1982.

Appendix D-4: Conversion of Manhours to Saskatchewan (Productivity Adjustment Factor)

A review of labour productivity based on petrochemical projects indicates that a multiplier of 1.25 be applied for conversion of U.S. Gulf Coast manhours to equivalent Canadian manhours.

For adjustment to Saskatchewan, the following two items must also be considered:

Unique Condition Factor

Projects of this size will generally experience during the construction period several occurances whereby labour will interpret certain management decisions as being an affront to their integrity and pride. It will be assumed that these types of non-strike situations will lead to a cumulative loss of one working day per month during the construction period.

Factor = one day/20 working days per month = 0.05

Climatic Factor

The Saskatchewan Region experiences considerably more sub-zero temperatures and the duration of the winter season is generally about two months longer than the Sarnia region. Therefore, an allowance of 20 days per winter will be made for adverse working conditions due to weather.

$$\text{Factor} = \frac{\text{20 days x 4 winters}}{\text{48 months x 20 days/winter}} = 0.08$$

Thus, the "Overall Productivity Adjustment Factor" is

= 1.25 x 1.05 x 1.08 = 1.417 = 1.42

Appendix D-5: Process Equipment and Plants Price Indexes

Year	Index	% Increase
1977	181.9	7.53
1978	195.6	10.70
1979	216.6	10.90
1980	240.2	11.25
1981	267.2	11.00
1982*	-	11.00

* Assumed that similar % incease will continue in 1982.

Source: Construction Price Statistics, "Chemical and Mineral Process Plant Price Indexes", Table 19, November 1981

Appendix D-6: Typical Manpower Distribution and Average Salary Scale

	% of Total Manpower *	Average Salary $ Month **
Operating Labour (OL)	34	1575
Operation Supervisors (20% of OL)	7	2500
Maintenance Labour (ML)	42	1875
Maintenance Supervision (20% of ML)	8.5	2500
Plant Management and Administrative Support	8.5	2700

Sources:

* K.A. Rogers and R.F. Hill, ESCOE, DOE, Report No. FE-2468-51, July 1979

** Values reported in SPC Study, (Job No. C8093) "Coronach Coal Gasification Study", Volume II, October 1977, have been adjusted to mid-1982 dollars using salary index in Appendix D-7.

Appendix D-7: Salary Increases for Operating Costs Calculation
(Location: Saskatoon)

Year	Index	% Increase
1976	204.0	10.25
1977	224.9	5.47
1978	237.2	6.62
1979	252.9	9.25
1980	276.3	7.38
1981	296.7	9.38
1982*	-	11.38

* Increase for 1982 assumed as 1981 + 2%

Source: Construction Price Statistics, "Union Wage Rate Index", Table 3, November 1981

Appendix D-8: Cost of Water Supply Calculated by SNC-Keith Gabriel, Saskatchewan

THE
KEITH GABRIEL
PARTNERSHIP
CONSULTING ENGINEERS AND ARCHITECTS

January 21, 1982 File: 6000

The SNC Group
1, Complexe Desjardins
P.O. Box 10
Desjardins Postal Station
Montreal, Quebec
H5B 1C8

Attention: Josef A. Mikhlin, ing., Ph.D.
Chef de Service, Carbon
Recherche et Developpement

Dear Dr. Mikhlin:

RE: Water Costs

The Saskatchewan Water Supply Board was contacted to obtain information regarding water rates charged to municipal and industrial clients.

The Saskatoon South-East Water Supply canal supplies water from Deifenbaker Lake to potash mines and communities along its route.

The Board charges their customers $1.10 for a 1000 imperial gallons of raw water. This price will likely be revised by10% to 20% in August, 1982.

I hope this information will be sufficient for your present requirements. Should additional information be required please don't hesitate to call.

THE KEITH GABRIEL PARTNERSHIP

Cliff A. Beattie, P. Eng.
Director of Business Development

CAB:bjg

[illegible] ANGUS STREET REGINA SASKATCHEWAN [illegible]
TELEPHONE (306) [illegible] TELEX [illegible]
A PARTNERSHIP BETWEEN KEITH CONSULTING ENGINEERS LTD
& GABRIEL ARCHITECTURAL GROUP LIMITED
REGINA
PRINCE ALBERT
EDMONTON
LETHBRIDGE

Volume

2

STUDY RELATING TO THE TECHNICAL AND ECONOMIC FEASIBILITY OF LIGNITE LIQUEFACTION IN SASKATCHEWAN

Volume II

ASSESSMENT OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES CONSIDERED FOR COMMERCIALIZATION BY 1990

Final Report
by

TECHNICAL DEVELOPMENT DEPARTMENT, SNC Inc

for

SASKATCHEWAN OIL AND GAS CORPORATION

and

CANMET, ENERGY, MINES AND RESOURCES, CANADA

October, 1982
Montreal

REQ. NO. 2-9203

STUDY RELATING TO THE TECHNICAL AND ECONOMIC FEASIBILITY OF LIGNITE LIQUEFACTION IN SASKATCHEWAN

CONTRACT REPORT NO. OSQ82-00269

Volume II

ASSESSMENT OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES CONSIDERED FOR COMMERCIALIZATION BY 1990

Final Report
by
TECHNICAL DEVELOPMENT DEPARTMENT, SNC Inc.
for
SASKATCHEWAN OIL AND GAS CORPORATION
and
CANMET, ENERGY, MINES AND RESOURCES, CANADA

Work on this project was conducted under the auspices of the Canada Centre for Mineral and Energy Technology, Energy, Mines and Resources, Canada

The project was funded 50/50 by Saskoil and CANMET, EMR, Canada

346+ pages

October, 1982
Montreal

MICROMEDIA

JUL 10 1984

PROJECT TEAM

STUDY MANAGER:	J.A. Mikhlin, Eng, Ph.D.
SENIOR PROCESS ENGINEERS:	F.A. Ashraf, Eng, M. Eng. J.A. Dhawan, Eng, Ph.D.
PROCESS ENGINEERS:	P. Khandhadia, Eng, M.S. J. Pienkowski, Eng, M.Sc.

Editorial assistance on the report was provided by Mr. S.F. Petzold.

ACKNOWLEDGEMENTS

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J.A. Mikhlin, Eng, Ph.D.

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1.0 INTRODUCTION

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1.0 INTRODUCTION

This study was carried out by the Coal Conversion Group, Technical Development Department of SNC Inc. to evaluate various coal liquefaction processing alternatives for two designated Saskatchewan lignites, i.e., from the Estevan and Willow Bunch deposits. The processes evaluated were subdivided into two categories.

- processes considered for commercialization by 1990, and
- processes considered for commercialization around 1995.

The scope of the work for the advanced processes (commercialization by 1990) included review and evaluation of direct and indirect coal liquefaction and pyrolysis processes listed in Table 1-1. The evaluation was made in two stages. Stage I comprises of technical assessment, prescreening and ranking of the processes for both Estevan and Willow Bunch lignites. Stage II covers techno-economic assessment of processes selected in Stage I but for Willow Bunch lignite only. The ultimate objective is to determine promising process(es) for Saskatchewan lignites.

The scope of the work in the second category (commercialization around 1995) is to ascertain state-of-the-art technology and review update of the processes and concepts listed in Table 1-2. A summary of the results is presented in Volume I.

In addition, an independant experimental program was carried out concurrently with this study to determine comparative reactivity of Estevan and Willow Bunch (Coronach) lignite. These tests were made by the Chemical Technology Division of B.C. Research and the results are presented in a separate report, "Liquefaction and Pyrolysis Tests on Samples of Estevan and Coronach Lignites", Project Report No. 2-02-691, March 1982. A summary of the results is presented in Volume I.

Table 1-1 Processes Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization by 1990

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
H-Coal	Hydrocarbon Research Inc.	U.S.A.
RAG/Veba	Ruhrkohle AG and Veba Oel AG	F.R.G.
Rheinbraun	Rheinische Braunkohlenwerke AG	F.R.G.
Exxon Donor Solvent (EDS)	Exxon Research and Engineering Company	U.S.A.
Liquid Solvent Extraction (LSE)	National Coal Board, Coal Research Establishment	U.K.
	COAL PYROLYSIS	
Lurgi-Ruhrgas	Lurgi Mineralöltechnik GmbH and Ruhrgas AG	F.R.G.
	INDIRECT COAL LIQUEFACTION	
SASOL-Type Fischer-Tropsch	SASOL Ltd.	South Africa
Mobil Methanol-to-Gasoline (M-MTG)	Mobil Research and Development Corporation	U.S.A.
	Gasifiers for Production of Synthesis Gas	
Lurgi (dry bottom)	Lurgi Mineralöltechnik AG	F.R.G.
British Gas/Lurgi (slagging)	British Gas Corporation and Lurgi Mineralöltechnik AG	U.K. F.R.G.
Koppers-Totzek	Krupp-Koppers GmbH	F.R.G.
Texaco	Texaco Research and Development Corporation	U.S.A.
Winkler (high temperature)	Rheinische Braunkohlenwerke AG	F.R.G.
Westinghouse	Westinghouse Electric Corporation	U.S.A.

Table 1-2 Processes and Concepts Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization around 1995

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
Zinc Halide	Consolidation Coal Co.	U.S.A.
Dow Coal Liquefaction	Dow Chemicals Co.	U.S.A.
Consol Synthetic Fuel (CFF)	Consolidation Coal Co.	U.S.A.
Solvent Refined Coal (SRC-1)	International Coal Refining Co.	U.S.A.
Solvent Refined Lignite (SRL)	University of North Dakota	U.S.A.
CO-Steam	Grand Forks Energy Technology Center	U.S.A.
Two-Step Liquefaction	Cities Service Co. C-E Lummus Co.	U.S.A.
	COAL PYROLYSIS	
Occidental Pyrolysis	Occidental Research Corporation	U.S.A.
	INDIRECT COAL LIQUEFACTION	
	Gasifiers for Production of Synthesis Gas	
Shell Gasifier	Shell International Petroleum	The Netherlands
U-Gas	Institute of Gas Technology	U.S.A.
	CO-PROCESSING	
Coil	Hydrocarbon Research Inc.	U.S.A.
Solvolysis	Mitsubishi Heavy Industries Ltd.	Japan
Veba Coal/Oel Co-processing	Veba Oel AG	F.R.G.

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1.1 OBJECTIVES OF THE STUDY

The main objectives of the study for advanced processes considered for commercialization by 1990 were identified as follows:

Stage I, Process Evaluation and Prescreening

- to determine state-of-the-art technology and predict behavior of Estevan and Willow Bunch lignite in the processes based on available data and previous experience of the process developers with similar coals.

- prescreening and ranking of the processes for further techno-economic assessment.

Stage II, Techno-Economic Assessment of Preselected Processes

- techno-economic assessment and comparison of pre-selected processes using Willow Bunch lignite as feedstock.

- selection of the promising process(es) for liquefaction of Willow Bunch lignite to produce synfuels substituting conventional transportation fuels.

- to determine constraints toward commercialization for selected process(es).

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1.2 TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION

Technical and economic criteria approved by the Saskatchewan Coal Liquefaction Technical Committee (the Study Committee) for direct and indirect liquefaction and pyrolysis processes are presented in detail in Volume I, "Summary and Process Comparison", however, a brief outline is given below.

1.2.1 **Feedstock Specifications**

Two Saskatchewan lignites from the Estevan and Willow Bunch deposits were designated as feedstocks for "prescreening" of coal liquefaction and pyrolysis processes (Stage I of the Study). Typical ultimate, proximate and petrographic analysis of these designated lignites are summarized in Appendix A. Both coals have high moisture and ash content. While average moisture content is the same for both coals, the ash content of Willow Bunch lignite and inertinite content are higher.

For techno-economic assessment of prescreened processes (Stage II of the Study), Willow Bunch lignite was designated as the feedstock. Detailed analyses of Willow Bunch lignite showing the "best" and "worst" coal properties are given in Appendix C.

1.2.2 **Product Criteria**

For prescreening of the processes (Stage I of the Study) the product criteria are:

- synthetic crude (naphtha, middle oil, heavy oil) via "direct" coal liquefaction.
- synthetic crude (naphtha, middle oil, heavy oil) via pyrolysis.
- gasoline as the main product via "indirect" coal liquefacation.
- pure methanol (99.5% purity) as transportation fuel or blendstock for gasoline via indirect coal liquefaction.

In order to allow a consistent comparison of the techno-economic assessments of preselected processes (Stage II of the Study), transportation fuels such as gasoline, diesel and jet fuel are considered as desired end products for both "direct" and "indirect" coal liquefaction processes. However, emphasis is given to production of gasoline. Light hydrocarbons (C_1-C_4) are considered as co-products.

1.2.3 Overall and Liquid Product Efficiency

In this study, a modified version of efficiency widely used for energy-related processes has been considered. The modified definition of "overall" product efficiency is the total gross-calorific-value (GCV) of all products divided by the total energy input of coal feed and electric power. For processes, where electricity is required from an outside source, a thermal equivalent factor of 10,000 Btu/kWh (10,550 kJ/kWh) as suggested by ESCOE*, rather than the true thermodynamic equivalent factor of 3413 Btu/kWh, has been used to allow realistic conversion loss for electric power generation.

Since the objective of the study is to investigate coal liquefaction processes for the production of transportation fuels, the "liquid" product efficiency is also calculated. Here, the liquid product refers to distillate product in the temperature range of C_5- 525°C.

1.2.4 Overall Plant Configuration, Plant Size and Battery Limits

All processes are considered as grass roots facilities. The plant comprises self-sufficient, integrated operation requiring coal, water and electricity (if required) as major raw material and process inputs.

The as-received coal supplied to the plant is assumed to be Run-of-Mine (ROM). No pretreatment/beneficiation of coal to reduce moisture and/or ash (minerals) contents is considered. Development of coal mine, water supply, gas delivery pipelines and booster stations, product transportation facilities, etc., are also not considered.

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* B.C. McBeath, The Engineering Society Commission on Energy Inc., "Guidelines for Coal Conversion Processes", DOE Report No.FE-2468-44, April 1979.

1.2 TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION (cont'd)

Plant Size and Capacity Factor : The conceptual, commercial plant will be based on a feedrate of 25,000 metric tons (tonnes) of as-received coal supplied to the plant. A capacity factor of 0.85, i.e., 310 working days per year is assumed.

Plant Location and Coal Delivery : The plant is assumed to be located within 16-24 km (10-15 miles) radius of the coal mines. No major site development other than leveling or grubbing has been assumed. Coal is assumed to be delivered to the plant by unit railcars. If, however, the plant site is within 3.0 km (2 miles) radius, transportation by trucks or conveyors may be economical.

Ash/Slag Disposal : Ash/slag is transported back to the mines for disposal.

Coal Fines : Coal fines may be burned for steam and power generation. Excess coal fines will be returned to the mines without any credits.

Water Availability : Water is assumed to be available to the plant site from Diefenbaker Lake through a canal to be built. Supply of water will also depend on Provincial Government priorities. It is further assumed that the cost of water will be the same as that for water presently supplied by the Saskatoon South-East Water Supply canal from Diefenbaker Lake to the potash mines and the communities along its route.

Steam and Power Availability : Steam is generated on site by recovering process heat and burning coal. Power is also generated onsite. Additional electricity, if required, will be purchased from the utility network and a conversion factor of 10,000 Btu/kWh (10,500 kJ/kWh) will be applied for thermal efficiency calculations.

1.2.5 Environmental Aspects

The conceptual plant design will meet all existing local, provincial and federal environmental regulations for particulates, liquid and gaseous effluents.

1.2.6 Economic Criteria

Preliminary or budget-type capital cost estimate, operating, maintenance and product cost evaluation is carried out for each of the "preselected" coal liquefaction processes only (Stage II of the Study).

Each process is evaluated as a self-sufficient, integrated, grass-roots facility processing 25,000 tonnes per day of as-received Willow Bunch lignite supplied to the plant. The plant is assumed to be located within 16 km (10 miles) of the coal mines in Saskatchewan. The service factor is assumed to be 85%, i.e., 310 operating days per year.

1.2 TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION (cont'd)

All costs are calculated in "constant" mid-1982 Canadian dollars and no forward escalation beyond mid-1982 included. Cost of capital is assumed to be 10% per annum. For 100% debt financing this value is equivalent to debt* cost in "real" terms. For 100% equity financing, it may be considered as return on equity before taxes.

Product costs for preselected coal liquefaction processes will be calculated on "Energy Value" and "Product Value" bases. The first method is based on the heating value only whereas the second also considers the quality of the individual product. The latter approach, Product Value method, was developed by The Engineering Society Commission on Energy Inc. (ESCOE)** and the objective is to allow relative comparison and ranking of the energy-related processes.

By-product credits have been taken considering \$65/t of sulfur, \$300/t of NH_3, and \$250/t of mixed phenols.

A large quantity of CO_2 is produced which has a potential for utilization in enhanced oil recovery operation. No credits are taken for CO_2 byproduct, since the market for such use is uncertain.

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* for calculations on "constant" dollar basis, debt cost is expressed in "real" terms, i.e., prevailing interest rates minus inflation.

** K.A. Roger and R.F. Hill, "Coal Conversion Comparison", DOE Report No. FE-2468-51, July 1979.

1.3 STUDY REPORT ORGANIZATION

The study report comprises of five volume:

Volume I	- Summary and Process Comparison
Volume II	- Assessment of Advanced Direct Coal Liquefaction and Pyrolysis Processess Considered for Commercialization by 1990
Volume III	- Assessment of Advanced Indirect Coal Liquefaction Processes Considered for Commercialization by 1990
Volume IV	- Review of Coal Liquefaction Technologies and Concepts Considered for Commercialization around 1995
Volume V	- Process Developer Data and Information

This volume II, "Assessment of Advanced Direct Coal Liquefaction and Pyrolysis Processes Considered for Commercialization by 1990", covers technical review and evaluation of five "direct" coal liquefaction and one pyrolysis process listed in Table 1-1 for both Estevan and Willow Bunch lignites (Stage I of the Study) and techno-economic assessment of the selected direct coal liquefaction process (Stage II of the Study) for Willow Bunch lignite only. The process comparison and prescreening results are presented in Volume I, "Summary and Process Comparison". The two-stage Liquid Solvent Extraction (LSE) direct coal liquefaction process was selected for techno-economic assessment.

1.3 STUDY REPORT ORGANIZATION (cont'd)

This techno-economic assessment of preselected LSE process is carried out for two extraction efficiencies considered for Willow Bunch lignite in the reactor (MAF basis):

- 65% extraction
- 82% extraction

The technical assessment of LSE process includes:

- Conceptual design of an overall plant configuration processing 25,000 t/d as-received Willow Bunch lignite, overall material balance, product efficiencies, and utilities summary.
- Product quality, analyses, and upgrading schemes.
- Environmental considerations.

The economic assessment of LSE process includes:

- Capital cost estimation and the total investment requirement for a 25,000 t/d plant located in Saskatchewan.
- Estimate of annual operating, maintenance and production cost.
- Product cost calculations using "Energy Value" and "Product Value" basis.

1.3 STUDY REPORT ORGANIZATION (cont'd)

This techno-economic assessment has been prepared based on information and data received from Coal Research Establishment, National Coal Board, England. The information package entitled "The Processing of Saskatchewan Lignite by the NCB's Liquid Solvent Extraction Process: Willow Bunch Lignite", (NCB Ref. No. DR496/2, 7/3, January 23, 1982) is included in the confidential Volume V, "Process Developer's Data and Information.

In addition, a sample of Willow Bunch lignite (Coronach #3216) was sent to NCB for preliminary characterization in the autoclave unit of the LSE process. Proximate, ultimate and petrographic analysis of the coal sample used in the experiments are presented in Appendix D. Complete NCB report on Willow Bunch lignite characterization is included in Volume V.

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2.0 REVIEW AND EVALUATION OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES

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2.0 REVIEW AND EVALUATION OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES

A review and evaluation of the following advanced (towards commercialization) direct coal liquefaction and pyrolysis processes is presented in this chapter. Comparative evaluation and prescreening results, as mentioned earlier, are presented in Volume I, "Summary and Process Comparison".

Process	Process Developer	Country
	DIRECT COAL LIQUEFACTION	
H-Coal	Hydrocarbon Research Inc.	U.S.A.
Exxon Donor Solvent (EDS)	Exxon Research and Engineering Company	U.S.A.
Modified I.G. Farben, New German Technology for High-Rank Coals	Ruhrkohle AG and Veba Oel AG (RAG/Veba)	Federal Republic of Germany
Modified I.G. Farben, New German Technology for Low-Rank Coals	Rheinische Braunkohlenwerke (Rheinbraun)	Federal Republic of Germany
Liquid Solvent Extraction (LSE)	National Coal Board, Coal Research Establishment	U.K.
	COAL PYROLYSIS	
Lurgi-Ruhrgas (L-R)	Lurgi Kohle und Mineraloltechnik GmbH	Federal Republic of Germany

2.0 REVIEW AND EVALUATION OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES (cont'd)

The coal liquefaction processes listed in the table have undergone various stages of development from microautoclave to bench-scale and process development units (PDU), and in some cases to the pilot plant scale. Microautoclave tests were conducted for preliminary characterization of various coals. The bench scale tests provided information on the effect of process variables such as temperature, pressure, coal feed rate, oil-to-coal ratio, etc.,on coal conversion and product yields. Evaluation of different catalysts (in catalytic processes), reactor modelling, cold-flow testing, process optimization studies, product analysis and upgrading were also part of the bench-scale programs.

The PDU runs demonstrated sustained and continuous operation and were primarily aimed at confirming the design basis and key operating conditions for the pilot plant. The pilot plant operation was intended to substantiate the feasibility of the processes and the mechanical realiability of equipment items, to verify the product yields and product quality on a large scale, and to obtain design and scale-up data for a large-scale commercial plant.

A generalized typical block diagram of an integrated, commercial, direct coal liquefaction plant is shown in Figure 2-1. The overall plant configuration and interrelationship between the main processing areas comprising the direct coal liquefaction plant are shown in this diagram.

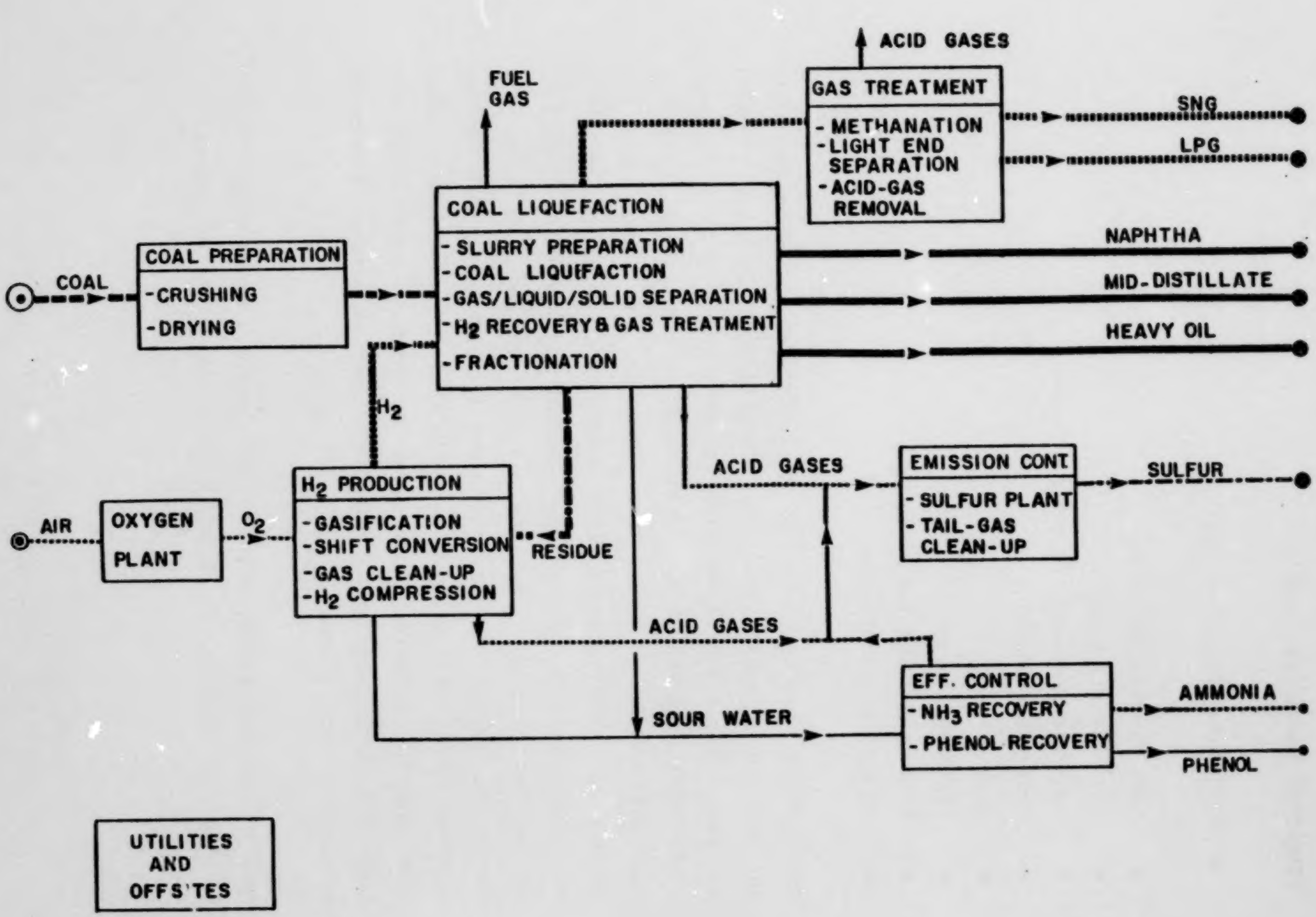

Figure 2-1 Typical Block Diagram of an Integrated Coal Liquefaction Commercial Plant

2.0 REVIEW AND EVALUATION OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES (cont'd)

The main processing areas are:

- Coal Preparation
- Coal Liquefaction
- Gas Treatment and Light-Ends Separation
- Hydrogen Production
- Oxygen Plant
- Emission Control
- Product Storage and Shipping
- Utilities and Offsites

The overall configuration of a conceptual, commercial plant includes coal preparation, coal liquefaction, product recovery and refining sections along with the required supporting systems, utilities supply and offsite facilities to comprise an integrated, self-sufficient operation. Coal and water are basic raw materials. Catalysts and other chemicals are also needed but in relatively smaller quantities. During steady-state operation, make-up hydrogen and fuel gas are produced by gasification of vaccum bottoms (or residues) obtained from the coal liquefaction section. Most of the steam is generated by recovering process heat and the balance is produced by burning coal fines and process-derived fuel gas. Electricity may be purchased from an external utility source or generated on site.

Most of the processing areas listed above (except coal liquefaction) are similar, in principle, for any of the processes considered in this volume. The coal liquefaction section is a unique area of the plant, characteric for a specific coal liquefaction process, and as such a coal-syncrude plant will differ mostly in the coal liquefaction section.

For instance, in a plant based on H-Coal process, which is a one-stage direct catalytic hydrogenation process, dried and pulverized coal is mixed with recycle coal-derived oil and the coal/oil slurry is fed together with hydrogen into high temperature and pressure ebullated-bed catalytic reactor (Co/Mo catalyst). In the RAG/Veba and Rheinbraun processes, the other two one-stage direct catalytic hydrogenation processes, a tubular up-flow reactor operating at high temperature and pressure is employed. Disposable red mud is added as a catalyst to the coal/oil slurry in the presence of molecular hydrogen.

In EDS and LSE processes, which are extraction processes (sometimes also referred to as "solvent refining" of coal), pulverized coal is slurried in a solvent and then the slurry reacts at elevated temperature and pressure. In the EDS process, the slurry oil is a "hydrogen donor" solvent which transfers hydrogen to the coal-oil in the reactor. The "donor" solvent is separately prepared by catalytic hydrogenation of recycle oil in a fixed-bed hydrotreater. The ability of the solvent to donate hydrogen to the coal-oil and stabilize the product is the key element of the EDS process.

In the LSE process, which is a two-stage direct coal liquefaction process, the coal is initially dissolved in an extractor (first stage reactor) and then after solid/liquid separation, the ash-free extract is fed together with molecular hydrogen into a fixed-bed hydrotreater (second stage reactor).

An integrated commercial plant based on a coal pyrolysis process may be presented by a block diagram similar to that shown in Figure 2-1. The main difference would be the substitution of the coal liquefaction section by a coal pyrolysis section.

Coal pyrolysis, also referred to as "devolatilization" or "carbonization", involves the heating of coal in absence of air or oxygen to yield heavy oil (tar), light liquids, gases and char. When coal is pyrolyzed, the hydrogen-rich volatile matter is released (devolatilization) and a carbon-rich solid residue or char is formed (carbonization). The composition and relative amounts of the products formed are influenced by the coal heat-up rate, temperature, residence time, coal particle size, hydrogen partial pressure, and reactor configuration. In the Lurgi-Ruhrgas process, fine coal particles are rapidly heated to the carbonization temperature by direct contact with hot recycle char to maximize the liquid yields. No hydrogen is added in the carbonization reactor. The liquid product (tar) is subsequently hydrogenated in a catalytic hydrotreater to produce distillate products.

A brief description of the similar processing areas of an integrated direct coal liquefaction or pyrolysis plant is presented below (except the coal liquefaction section which is described in more detail for energy process separately):

Coal Preparation

This section includes coal handling and storage facilities along with grinding, classifying and drying equipment to provide coal of desired specifications. Raw coal from the mines is unloaded and conveyed to open pile storage areas.

Bucket-wheel reclaimers remove the coal from the storage pile onto a system of belt conveyors from where it is distributed to coal storage bins. During conveying, coal is sampled and iron present in it is removed by magnetic separation methods.

The raw coal from the coal storage bins is fed into a crusher/dryer where it is simultaneously pulverized and dried to the desired specification. Hot combustion gas used for drying the coal also conveys the coal to cyclone classifiers for removal of dust and fines. The dried coal from the bottom of the cyclones is delivered to coal storage bunkers from where it is conveyed to the coal liquefaction section.

Gas Treatment and Light-End Separation System

The gases from the coal liquefaction section are further processed in the gas treatment and light-ends recovery section.

Gas Treatment and Light-End Separation System (cont'd)

High-pressure sour gases containing unreacted hydrogen are initially treated in an amine scrubber to remove the acid gases. The treated gas is then cryogenically purified to produce a hydrogen-rich stream which is recycled to the coal liquefaction reactor (in H-Coal, RAG/Veba, and Rheinbraun processes), to the solvent hydrotreater (in EDS process) or to extract/tar hydrogenation reactor (in LSE and L-R processes). Acid gases from gas treatment are sent to an emission control unit for sulfur recovery. The remaining treated gases (sweet gas) containing mainly lighter hydrocarbons (C_1-C_4) are further processed in the gas plant where the valuable products, i.e. propane and butane (LPG) are separated by conventional petroleum technology.

Hydrogen production

The residual slurry (or solid residual, char in the case of the LSE and L-R processes) leaving the coal liquefaction section is gasified with oxygen and steam to produce synthesis gas. The gaseous product mixture consisting mainly of H_2, CO, CO_2, H_2O and H_2S undergoes a subsequent shift-conversion step where carbon monoxide and steam are converted to hydrogen and carbon dioxide via the water-gas reaction. The shifted gas is then treated in order to selectively remove H_2S and CO_2 from the hydrogen-rich stream. The H_2S-free carbon dioxide stream is discharged to the atmosphere and the H_2S-rich gas stream is sent to a Claus plant for sulfur recovery. The purified hydrogen-rich stream is compressed and sent to the coal liquefaction section.

Oxygen Plant

High-purity oxygen (99.0 mol%) required for the gasification of the vacuum bottoms (or residual char) is produced in a conventional cryogenic air separation unit. At the same time, nitrogen for stripping in the acid gas removal unit and for use as an inert gas in the storage of liquid hydrocarbon products is also obtained.

Emission Control Section

The acid gas streams from various processing sections are combined and then processed in a conventional two-stage Claus plant in order to convert H_2S to elemental sulfur and to produce an offgas stream suitable for discharge to the atmosphere. Elemental sulfur is produced in a molten form which can be converted to flakes for shipment.

Effluent Control Section

Sour water collected from different sections of the plant contains crude-mixed phenols, and dissolved NH_3, H_2S, and CO_2 as impurities. Dissolved gases are removed from the sour water by stripping and then processed for recovery of ammonia. An ammonia stream essentially free of H_2S is produced and sent to storage, whereas the acid gases are sent to the Claus sulfur plant for recovery of sulfur. The phenol-containing water from sour water

Effluent Control Section (cont'd)

stripping is treated in a dephenolization process for removal of phenols from the aqueous stream. A mixture of phenols, often referred to as crude-mixed phenols is produced and sent to storage.

Utilities and Offsites

Utilities service include a full range of systems necessary for the operation of the plant, such as steam generation, electric power and fuel-gas distribution, cooling water, boiler feed water and waste water treatment units. Most of the steam required for the plant is generated by recovering process heat. Additional high-pressure steam is produced by coal-fired steam boilers which are also used during the start-up operation. Superheated steam is provided by a fuel-gas fired steam superheaters. The plant's water system includes a raw water reservoir, cooling towers and demineralizers for boiler feed water. A neutralizing sump is also provided for biological and waste water treatment.

Offsite facilities include administration buildings, laboratory, maintenance equipment and buildings, warehouses, a firehouse, roads and onsite railway trackage.

Having outlined the similar processing areas of a typical coal liquefaction plant, a detailed description of the charactericstic part of the process, the coal liquefaction section, is given

2.0 REVIEW AND EVALUATION OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES (cont'd)

in the technical review of each process. The technical review presented in the following pages includes history and development of each process with an overview of bench-scale, PDU and pilot plant programs. The scale of operations, time period of investigation, range of operating conditions investigated for a particular type and rank of coal (s) with some pertinent results are also presented.

The experimental results of bench-scale, PDU, and pilot plant runs presented in this study for various processes are taken from different sources. Data reported by the process developers are somewhat inconsistent in the sense that in some cases product yields are reported on MF basis whereas other have preferred to use MAF basis. In such a case, data have been converted to MAF basis, where possible.

Another source of inconsistency in available literature data is the cut-points for various distillate fractions. For example, some process developers reported as naphtha as hydrocarbons in the boiling range (C_5-200°C), mid-distillate as (200-325°C), heavy oil as (325-525°C) and residual oil as (>525°C). In actuality some overlaps exist and the cut-points may be something like naphtha (C_4-205°C), mid-distillate (175-345°C), heavy oil (260-525°C), and the residual oil (>525°C). However, the amount of liquid present in the overlap region is usually small, for example C_4 in H-Coal naphtha fraction is less than 1%.

2.0 REVIEW AND EVALUATION OF ADVANCED DIRECT COAL LIQUEFACTION AND PYROLYSIS PROCESSES (cont'd)

In this study, the cut-points are presented without an overlap in the temperature range. Overlap in the individual distillate fraction is neglected since the desired product for prescreening of "direct" coal liquefaction processes, as set out by the Study Committee, is coal-derived syncrude (C_5-525°C). Also, the cut-points given in Fahrenheit units by some process developers have been converted to Centigrade units and rounded off to the nearest multiple of ten, for example (C_5-400°F) has been taken as (C_5-200°C).

In the technical review, coal(s) which have been extensively tested by the process developer and/or those similar to the designated Saskatchewan lignites (Estevan and Willow Bunch) have been identified. Comparison of these coals have been made and the significance of their analytical similarities or differences on product yields and efficiencies for the Saskatchewan lignites have been discussed.

The Saskatchewan lignites have inherently high moisture and ash content which affect their processibility in a coal liquefaction process. For instance, moisture of the feed coal must be reduced to a desired level (2-4 wt%) before coal is fed into the reactor. However, drying of low-rank coals to such an extent with hot air has been observed to reduce activity of the coal with respect to liquefaction, possibly because of surface oxidation and collapse of pore structure. On the other hand, if excess moisture is not removed then it may be necessary to increase necessary to increase the total operating pressure inside the reactor to maintain a given partial pressure of hydrogen. Also, presence of moisture creates undesirable oil/water emulsion and foaming problems in the coal/oil slurry preparation unit.

The technical review includes an estimate of product yields and efficiencies for a conceptual, commercial, self-sufficient plant processing 25,000 t/d of as-received coal. Such calculations were made for coal(s) widely tested by the process developer and/or similar coal(s) to the designated Saskatchewan lignites. Similar calculations have been made for Saskatchewan lignites and comparison of the product yields and efficiencies with the "reference" coals have been made. A discussion on product quality and its analysis has been presented to the extent data were available.

Potential problem areas pertaining to the processing of Saskatchewan lignites and other general areas requiring further development have been identified. The areas of technical concern are both technology-related and hardware-type such as slurry mixing and pumping, preheater furnace, reactor control and performance, slurry handling, pressure letdown valves, instrumentation, materials of construction, environmental impact, wastewater treatment, occupational health and safety. In some processes, the problems are generic in nature whereas in others, further improvement in reliability of operation and less maintenance is required.

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2.1 H-COAL PROCESS

H-Coal is a trade name of a patented catalytic hydroliquefaction process (U.S. Patent 3,321,393; Canada Patent 965,720) developed by Hydrocarbon Research Inc. (HRI) for the conversion of coals to produce mainly boiler fuels and/or synthetic crude oil.

In the H-Coal process, coal is pulverized, dried, and then mixed with recycle oil to form a coal/oil slurry containing approximately 50% solids by weight. The mixture is preheated and fed into an ebullated-bed catalytic reactor where the coal is liquefied at high temperature (425-455°C) and pressure (15-21 MPa) in the presence of hydrogen. The reactor effluents leaving from the top of the reactor are cooled and then separated into gases, liquids and a residue slurry by conventional flash separation techniques. The catalyst being heavier than the products is retained in the reactor. The spent catalyst is periodically withdrawn and replaced by fresh catalyst through the on-line "catalyst addition and withdrawal system". The primary products from the process are LPG, naphtha (C_5-200°C), mid-distillate (200-345°C) and heavy oil (345-525°C). By-products include sulfur, ammonia and crude phenols.

The novelty of the H-Coal process is the application of the commercially proven ebullated-bed reactor, originally developed by HRI for the H-Oil process to convert heavy-oil residues into lighter fractions. Potential advantages of the catalytic, ebullated-bed reactor and other main features of the H-Coal process are:

- Coal of any rank or type may be processed. No pretreatment to alleviate caking tendency or calcium carbonate deposition is required.

- High liquid (C_5-525°C) yields (45 to 50 wt% on MF coal basis or about 50 to 56 wt% on MAF coal basis) are achieved, and the conversion is not dependent on catalytic effect of the ash.

- Continuous catalyst replacement controls deactivation, provides constant product quality, and allows the possibility of continuous catalyst regeneration.

- Operating conditions can be varied to produce the desired product yields, i.e. from heavy boiler fuel as main liquid product (fuel-oil mode under less severe operating conditions, reaction space velocity of 1120 $kg/h/m^3$) to all distillate synthetic crude oil comprising of lighter hydrocarbon fractions (syncrude mode under more severe operating conditions, space velocity of 480 $kg/h/m^3$).

2.1.1 History and Development of the Process

The H-Coal process has been under development at HRI since 1964. More than fourteen types of coal have been evaluated in over 68,000 hours of operation. The development and demonstration of the process have been carried out in micro-autoclaves, bench-scale units (11 kilograms per day of coal) and a process development unit (PDU) having a capacity of processing up to 3.2 tonnes per day of coal. Recently, the feasibility of the process was being demonstrated in a pilot plant located near Catlettsburg, Kentucky. The pilot plant has a capacity of processing up to 225 tonnes per day of coal in the syncrude mode and up to 545 tonnes per day of coal in the fuel-oil mode.

A detailed design study for a commercial H-Coal plant to be located at Breckinridge County in Kentucky was initiated under the sponsorship of U.S. Department of Energy as a cooperative venture with Airco Energy Co. and Ashland Oil Inc. A proposal for U.S. government support has been submitted to the U.S. Synthetic Fuels Corporation. The current schedule calls for detailed engineering design to begin in 1982 leading to construction in 1983 with initial production in 1988.

The present H-Coal process development program consists of laboratory studies, product testing and upgrading studies, and the operation of the pilot plant which has been funded up to 1982. The total cost of the program (1974-82) is expected to reach 320 million. The sponsors of the H-Coal program are the U.S. Department of Energy, Electric Power Research Institute, Ashland Oil Inc., Standard Oil of Indiana, Conoco Coal Development, Mobil Oil Corp., the Commonwealth of Kentucky, and Ruhrkohle AG.

2.1.1.1 Bench-Scale and PDU Program

Figure 2-2 is a simplified diagram of a bench-scale unit capable of processing up to 11 kg/d of coal. In a typical bench-scale experimental run, feed coal of minus 0.25-mm size is slurried with recycle oil in an oil-to-coal ratio typically between 2 and 3 to 1. The slurry is then pumped to a reactor pressure between 13 to 17 MPa through a feed pump. The feed slurry is mixed with hot internal recycle slurry from the preheater. Make-up hydrogen makes one pass through the preheater and is mixed with the internal recycle; it then flows through the lead bath preheater which is held in the temperature range of typically 405-415°C.

2.1 H-COAL PROCESS (cont'd)

The preheated hydrogen/recycle-slurry mixture is mixed with the feed slurry and fed into the bottom of an ebullated-bed reactor. The reactor, 2 cm in diameter and 3 m long, is surrounded by an electrically heated molten-lead bath to maintain the selected reaction temperature (455°C).

The internal recycle stream is separated from the vapors and is recycled through the hot oil recycle pump. The net effluents are cooled and separated into gases, liquids and slurry streams. A portion of the slurry (separator bottoms) is recycled directly to the coal/oil preparation tank and the remainder is collected to be fed into either one of the two oil-recovery-system*: In each case, the net products are collected separately, weighed (or metered) and analyzed according to standard procedures.

The 3.2-t/d PDU, while it is functionally similar to the bench-scale unit described above, is more elaborate and simulates details of the Catlettsburg pilot plant. During the PDU program, several modifications have been made for improved and reliable operation. Some of these include: a modified hot-slurry feed system, redesign of reactor internals, installation of an "on-line catalyst addition and withdrawal" system, and an addition of a hydroclone system for continuous solid/liquid separation.

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* Not shown on Figure 2-2

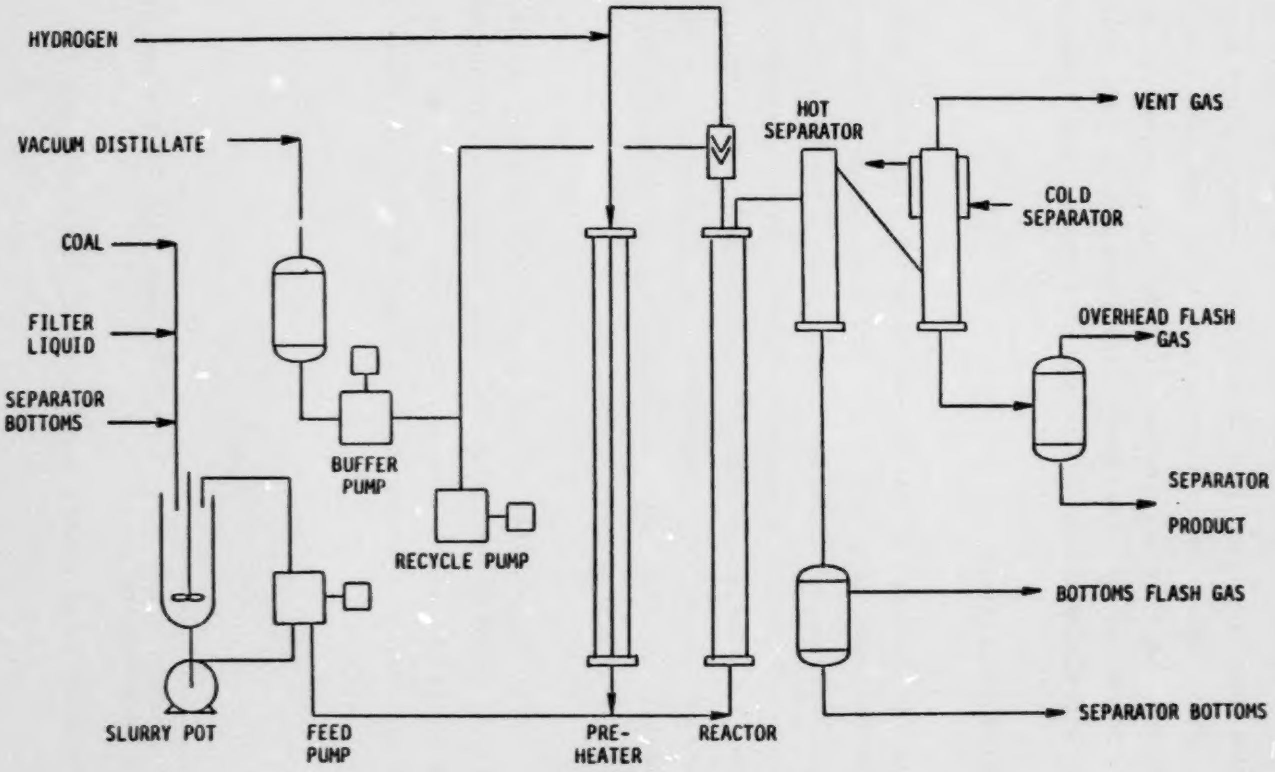

Figure 2-2 Schematic Diagram of the H-Coal Bench-Scale Unit (11 kg/d of coal)

Source: Proceedings of the EPRI Contractor's Conference on Coal Liquefaction, May-8-10, 1979

Over 60,000 hours of bench-scale operation and 8,000 hours of PDU operation have been achieved. In addition, several other supporting studies and process improvement investigations have been carried out. A complete review would be quite extensive and is beyond the scope of this study, therefore, only some relevant results pertaining to the processing of lignites are presented.

- **Testing of Lignites**

Although the process is flexible in terms of coal feed, only two experimental runs have been made with lignites on a bench-scale level*. In 1966, North Dakota lignite** from the Dickinson area was tested (550 hours of operation) when the program was under the sponsorship of the Office of Coal Research, U.S. Department of Interior (now U.S. Department of Energy). A second experimental run (705 hours of operation) was made in 1967 with Texas lignite for a private client.

Typical operating conditions and normalized product yields for North Dakota lignite are given in Table 2-1. The evaluation of this coal was carried out at two feed rates (space velocity 300 and 500 $kg/h/m^3$) at a temperature of about 450°C. The results

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* Some microautoclave tests have been made with Australian brown coals (see Section 2.1.3.1).

** Analysis of North Dakota lignite is given in Appendix B.

Table 2-1 Product Yield and Distribution for North Dakota Lignite Obtained in H-Coal Bench-Scale Experimental Runs

		HRI Run No. 177-16 (132-168 h)		HRI Run No. 177-16 (335-382 h)	
Reactor Temperature, °C		450		451	
Pressure at Reactor Outlet, MPa		15.4		15.5	
Coal Feed to Reactor					
Moisture, wt %		2.3		9.4	
Ash, wt % MF coal		9.7		11.1	
Space Velocity, kg/h/m^3		300		500	
Product Yields, wt% coal		MF	MAF	MF	MAF
Hydrocarbon Gases		15.1	16.7	16.5	18.5
Naphtha	(C_5-200°C)	17.9	19.8	17.8	19.9
Middle Oil	(200-345°C)	9.9	10.9	4.1	4.6
Heavy Oil	(345-525°C)	3.5	3.9	2.3	2.6
Residual Oil	(> 525°C)	7.9	8.7	11.3	12.6
		54.3	60.0	52.0	58.2
Unconverted Coal (ash free)		14.6	16.1	16.9	18.9
Ash		9.6	-	10.7	-
CO and CO_2		7.5	8.3	11.5	12.9
H_2O		18.5	20.5	12.6	14.1
H_2S		1.0	1.1	0.7	0.8
		105.5	106.0	104.4	104.9
Hydrogen Consumed		5.5	6.0	4.4	4.9
		100.0	100.0	100.0	100.0

Source: HRI Report, "Project H-Coal", prepared for Office of Coal Research, U.S. Department of Interior, Contract No. 14-01-0001-477, 1967 indicate that 52 to 54 wt% of MF coal (or 58 to 60 wt% MAF coal) is converted to hydrocarbon gases and liquids. Total conversion of coal is in the range of 82 to 85 wt%.

- **Low-Pressure Runs**

A bench-scale run (HRI Run No. 177-138) on Illinois No. 6 coal successfully demonstrated operations at a hydrogen partial pressure of 7.8 MPa. This corresponds to a total pressure of about 14 MPa in the reactor as compared to the 17-21 MPa pressure range normally employed in the PDU and pilot plant operations.

The total liquid yield decreased by 5 wt% for this low-pressure run as compared to the previous 21 MPa operation. However, lower design pressure may result in lower capital and operating costs and this alternative will be further explored by HRI in their future studies.

- **Effect of Ash Content of Coal on Product Yield**

A run-of-mine Illinois coal containing 18 wt% ash and a deep-cleaned Kentucky No.9 with 6.4 wt% ash were processed under similar conditions. Conversion for the deep-cleaned coal was 3.7 wt% higher on MAF basis. The distillate yield (C_5-525°C) was 3.4 bbl/t of MAF Kentucky coal versus 3.3 bbl/t MAF Illinois coal. This indicates that lower ash content in the feed coal favors higher conversion of coal on MAF basis and higher distillate product yields.

- **Testing of Improved Catalysts**

A series of bench-scale tests (HRI Run No. 177-130 to 177-137) was made to identify a better catalyst which can lead to a higher distillate yield, reduced nitrogen and sulfur contents in the product, and having improved regeneration characteristics. Four new catalysts were compared with the standard cobalt molybdate catalyst (American Cynamid 1442A) in similar operating conditions using bituminous Illinois No. 6 coal which has high sulfur (3.6 wt% MF coal) and nitrogen (1.2 wt% MF coal) contents. The catalysts used by HRI are:

Catalyst	Composition	Bulk Density (g/cm^3)
• American Cynamid 1442A	$CoMo/Al_2O_3$	0.580
• Amocat 1A	$CoMo/Al_2O_3$	0.675
• American Cynamid SN 4475	$NiMo/Al_2O_3$	0.585
• Armak PA 24268	$NiMo/Al_2O_3$	0.802
• Amocat 1B	Mo/Al_2O_3	0.714

These catalyst evaluation tests were carried out at an intermediate space velocity (800 $kg/h/m^3$) so that the results could be projected to the fuel-oil mode (1120 $kg/h/m^3$) or syncrude mode (480 $kg/h/m^3$) of operation. Other process conditions were:

• Coal Feed Rate	0.8 kg/h
• Slurry Oil Rate	2.0 kg/kg of coal
• Reactor Temperature	455°C
• Pressure	15.5 MPa
• Hydrogen Partial Pressure	12.4 MPa at Reactor Outlet

Of the catalyst tested, Amocat 1A gave the best overall performance. This catalyst is a cobalt molybdate on alumina with a pore size distribution maximized at 125 Å (12.5 x 10^{-6}mm) pore diameter instead of 55 Å for the "base" catalyst (American Cynamid 1442A). When compared to the base catalyst, the distillate yield increased by 3 wt% on MF basis, residual oil yield decreased by 5.4 wt% on MF basis (from 26 wt% to 20.6 wt%), and the residual-oil sulfur content reduced by 0.08 wt% from a level of 0.45 wt% on MF basis. In kinetic terms, the catalyst was 20% more active than the "base" catalyst for secondary conversion of residual oil to lighter distillate products and 40% more effective in desulfurization of the residual oil. The nitrogen content of the residual oil was nearly the same as with the base catalyst but the viscosity was reduced to half. Most of the original pores were recovered after regeneration.

American Cynamid SN 4475 was the next best catalyst in overall performance. The reduction in the yield and sulfur content of residual oil was about 65% of the improvement obtained with the Amocat 1A catalyst. The nitrogen content was lower by about 0.1 wt% than that obtained with Amocat 1A, particularly in the lighter distillate products.

Armak PA 24268 gave the best performance in terms of the nitrogen content of the liquid products. The nitrogen content of naphtha was about one-third as compared to that obtained from the base catalyst run.

Amocat 1B produced less residual oil than the base catalyst and about the same as the SN 4475 catalyst. Compared to base catalyst, the nitrogen content of the product was about the same but the sulfur content was slightly lower. Recovery of catalyst pores was quite low (62-80%) compared to about 90% for the other catalysts.

- **PDU Runs and Results**

Thirteen PDU runs, each typically of 30-day duration, have been carried out with three selected coals, namely bituminous Illinois No.6, Kentucky No.11 and subbituminous Wyodak coals.

Of the PDU runs made so far, PDU Runs No. 5 and No. 6 using the "base" catalyst are of particular importance. Typical oper-ating conditions and product yields for these runs are shown in Table 2-2.

Prior to PDU Run No.5, the PDU was modified to permit operation of an on-line "catalyst addition and removal" system. Consequently, PDU Run No.5 was designed to demonstrate the syncrude mode of operation at equilibrium catalyst condition using Illinois No.6 coal. The run lasted for 30 days as planned at steady-state catalyst activity without any significant operational problems.

PDU Run No.6 was performed in the syncrude mode using subbituminous Wyodak coal which has a high calcium (2.0 wt% MF coal) and oxygen (18 wt% MF coal) content. Coal conversion of 91 to 93.5 wt% on MAF basis instead of 90%*, distillate yield of 48 wt%

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* Values predicted for PDU runs on the basis of bench-scale experiments.

MAF instead of 51 wt%* and gas yield of 15.3 wt% MAF instead of 12.3 wt%* were achieved. Hydrogen consumption was higher, 6.8 wt% MAF instead of a projected 5.7 wt%*, reflecting that less severe operating conditions for this coal may be required. The run operated for thirty days, as planned, despite two interruptions due to carry over of catalyst into the pressure letdown valves and incomplete drying of the feed coal. To prevent catalyst carry over, a higher density catalyst were explored. Also, additional drying equipment was installed for high-moisture content (30 wt%) Wyodak coals. No calcium-induced solid deposits were found inside the reactor, a problem which has been reported by other process developers (EDS and SRC)**.

PDU Run No.10 was in the syncrude mode with Wyodak coal using one of the new catalyst (not disclosed by HRI) previously tested in the bench-scale unit. The data on this run is not available but as per communication from HRI***, the product yield and distribution were nearly the same as those predicted on the basis of bench-scale experiments. As compared to the standard cobalt molybdate catalyst, operation with this new catalyst resulted in lower hydrogen consumption.

- **Other Process Related Studies**

Cold-flow models to simulate the actual bench-scale, PDU, and pilot plant reactors have been constructed at R&D Centre of HRI at Lawrenceville, New Jersey. The cold-flow tests were carried out to investigate the fluid dynamics of the ebullated-bed

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* Values predicted for PDU runs on the basis of bench-scale experiments.

** Exxon Donor Solvent process, Exxon Research and Engineering. Solvent Refined Coal process, Gulf Oil Corporation, U.S.A.

***J.L. Gendler, HRI, Personal Communication to J.A. Mikhlin, SNC Inc., Letter No. 01-JLG-17-81, Jan. 23, 1981.

Table 2-2 Product Yield and Distribution for PDU Run No.5 (Bituminous Illinois No.6) and PDU Run No.6 (Subbituminous Wyodak)

		PDU Run No.5 (Period 29)		PDU Run No.6 (Period 17)	
Temperature, °C					
Coal Preheater Outlet		394		323	
Reactor		455		448	
Pressure, MPa					
Hydrogen Partial Pressure at Reactor Outlet		13.2		11.2	
Coal Feed to Reactor					
Space Velocity, $kg/h/m^3$		504		482	
Moisture, wt %		3.0		18.2*	
Ash, wt % of MF coal		10.8		9.3	
Catalyst Added, Cummulative, kg/t MAF coal		0.6		0.6	
Product Yields, wt% coal		MF	MAF	MF	MAF
Hydrocarbon Gases	(C_1-C_4)	13.05	14.75	14.33	15.53
Naphtha	(C_5-200°C)	16.27	18.39	21.74	23.56
Middle Oil	(200-345°C)	20.66	23.35	13.86	15.02
Heavy Oil	(345-525°C)	7.72	8.72	8.41	9.12
Residual Oil	(> 525°C)	19.00	21.47	11.46	12.42
		76.70	86.68	69.80	75.65
CO		0.16	0.18	0.44	0.48
CO_2		0.37	0.42	2.46	2.67
H_2S		2.64	2.98	0.87	0.94
NH_3		1.08	1.22	0.75	0.81
H_2O		6.65	7.51	17.13	18.57
Unconverted Coal		5.78	6.53	7.09	7.68
Ash		11.51	-	7.73	-
		104.89	105.52	106.27	106.80
Hydrogen Consumed		4.89	5.52	6.27	6.80
		100.00	100.00	100.00	100.00

* Incomplete drying

Source: EPRI Report AF-1143-SR, July 1979 reactor under different operating conditions and to obtain information related to the scale-up aspects of the reactor system.

Different methods of solid/liquid separation such as magnetic separation, centrifugation, filtration, solvent precipitation and hydrocyclone systems have been investigated using small-scale models of commercial-type equipment. Hydroclone separation for the syncrude mode and solvent precipitation for the fuel oil mode were selected and have been incorporated in the pilot plant for further testing.

In the H-Coal process, the pulverized coal is normally dried to about 2 to 4 wt% moisture before it is fed into the coal/oil slurry preparation tank. In some experimental tests, a higher moisture content of 9 to 20 wt% due to insufficient drying resulted in foaming problems and created undesirable oil/water emulsions in the product recovery vessels. It was found that raising the temperature inside the coal/oil slurry preparation tank from 90-105°C to 130-150°C resulted in the removal of excess moisture without any foaming or emulsion problems.

2.1.1.2 Pilot Plant Program

The H-Coal pilot plant is located on a 20-hectare site adjacent to the Big Sandy River in Boyd County, Kentucky, between Catlettsburg and the Ashland Petroleum refinery. The plant has a design capacity of processing up to 545 tonnes per day of coal in the fuel oil mode or 225 tonnes per day of coal in the syncrude mode. Most of the process design and engineering of the pilot

2.1 H-COAL PROCESS (cont'd)

plant, except for the Lummus anti-solvent deashing system for solid/liquid separation in the fuel oil mode, was carried out by HRI. Construction of the plant was done by Badger Plant Inc. and the operation is being carried out by Ashland Synthetic Fuels Inc.

A simplified flow diagram of the H-Coal pilot plant is given in Figure 2-3. Since the existing pilot plant is still too small for complete, integrated, self-sufficient operation, some of the facilities have been tied up with the nearby Ashland petroleum refinery. Pipelines from the refinery supply hydrogen, fuel gas, propane, process and boiler feed water, and instrument air to the pilot plant. Furthermore, sulfur containing gases from the acid-gas scrubber and sour-water system are sent to the refinery for further processing and recovery of elemental sulfur in a Claus plant. All other facilities such as steam generation, fuel distribution system and waste water treatment are incorporated in the existing pilot plant. In addition to this, measures have been taken to demonstrate the compliance of the process with EPA regulations. These regulations require crushing, grinding and handling of coal in closed environment, installation of filter bags and cyclones to trap particulate emissions, scrubbers to remove gaseous impurities and treatment of sour water before it is discharged to the nearby river.

Most of the oil products is sent to the adjacent Ashland refinery and the remainder is being stored offsite to povide feedstock for future testing.

As shown in Figure 2-3, the pilot plant has been designed for two modes of operation, i.e. syncrude and fuel oil. In the syncrude mode, the slurry leaving the steam stripper is sent to a vacuum distillation tower to recover additional light oil whereas in the fuel oil mode the slurry passes to a Lummus anti-solvent deashing step (shown by dotted lines) to separate the heavy boiler-fuel from the residual slurry. In both cases, the residual slurry containing some heavy oil, unreacted coal and ash is sent to a flaker to convert the residue material into solids which are presently being disposed as land fill. Some residual slurry is being stored for future tests. In commercial plant operation, the residual slurry will be gasified to generate make-up hydrogen for the process. Although the pilot plant will not demonstrate this step, about 100 tonnes of residual slurry from HRI's process development unit has been tested by Texaco in their Montebello pilot plant with satisfactory results.

The operating program (1980-82) of the Catlettsburg pilot plant includes processing of bituminous Kentucky No.11, Illinois No.6 and subbituminous Wyodak coals. According to the previous schedule, testing of Kentucky No. 11 and Illinois No.6 coals in both modes of operation, and Wyodak coal in the syncrude mode was planned. Due to delay in the program and shortage of funds, the operation of the pilot plant in the fuel-oil mode has been suspended, and testing is being carried out in the syncrude mode only.

The pilot plant was commissioned in February 1980 and has completed a series of scheduled oil runs with light gas-oil, heavy gas-oil and residual fuel oil. These dry run (i.e. without coal

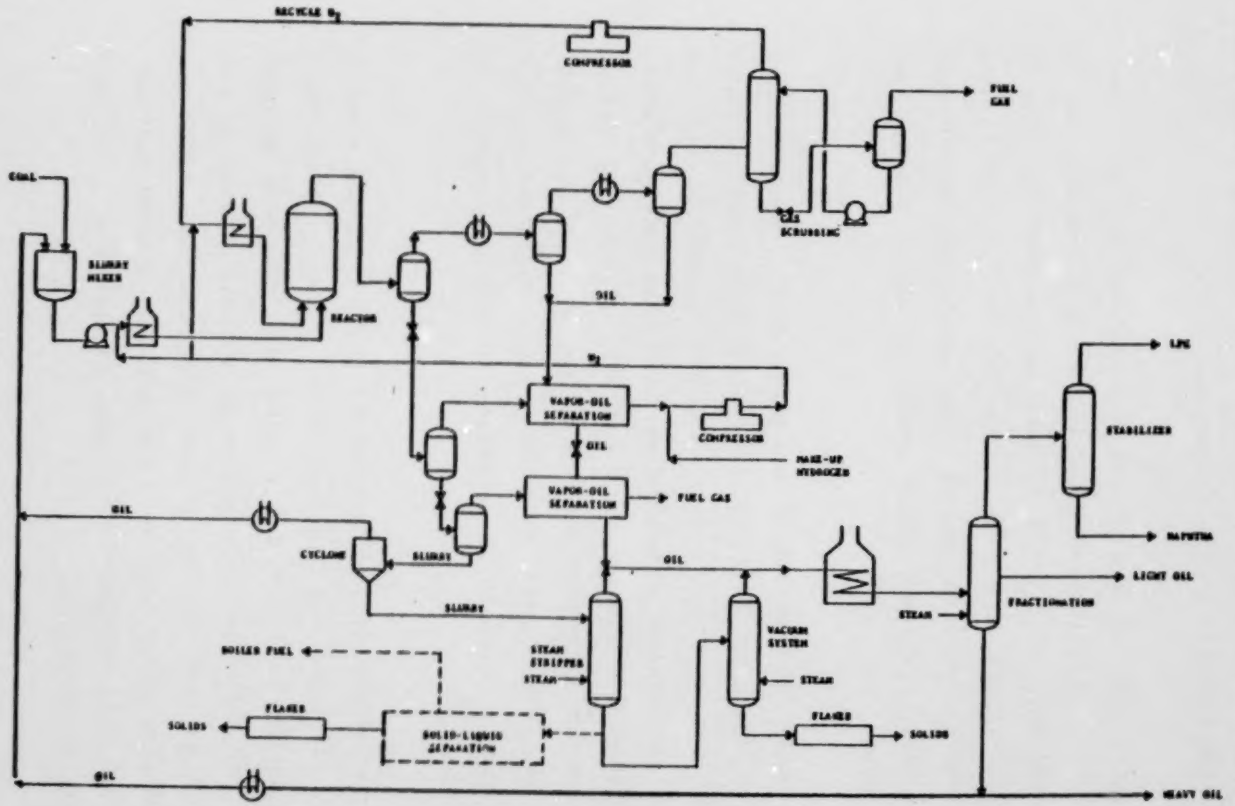

Figure 2-3 A Simplified Process Flow Diagram of the H-Coal Pilot Plant at Catlettsburg, Kentucky

Source: H.H. Stotler and R.T. Schutter, "Status and Plans of H-Coal Pilot Plant" Coal Processing Technology, Vol.5, 1979.

feed) were made to check the operation of the equipment items and also to provide training to the operators.

Coal was first fed to the H-Coal reactor on May 29, 1980. Initial break-in operations were carried out with Kentucky No. 11 coal at a feed rate of 200 t/d of coal in the syncrude mode of operation. On-stream days were limited due to frequent failure of the pressure let-down valves and unexpected maintenance requirements for equipment items such as slurry pumps, heat exchangers, etc.

Following a major turn-around of equipment items, the operation was initiated again in February 1981 with bituminous Illinois No.6 coal in the syncrude mode. The run lasted 45 days as planned and continuous, trouble-free operation was achieved. Table 2-3 presents a summary of the H-Coal pilot-plant design product yields for different coal feeds and modes of operation. Tables 2-4 and 2-5 present a typical day's operation in the 45-day run showing the actual operating conditions compared to the target values.

The third experimental run, with Kentucky No.11 coal was started on August 7, 1981 in the syncrude mode. Some problems were encountered during this run due to the high chloride content (0.3 wt% MF) of this coal. Corrosion in the ball valves of the slurry feed-pumps and in the product pipelines was observed. Finally, a massive power failure in the adjacent Ashland refinery resulted in the loss of make-up hydrogen to the pilot plant and the run had to be terminated.

Table 2-3 Design Product Yields for the H-Coal Pilot Plant

Coal		Bituminous Illinois No.6 (Burning Star Mine)		Subbituminous Wyodak
Mode of Operation		Syncrude	Fuel Oil	Syncrude
Product Yields, wt % MF coal				
Hydrocarbon Gases	(C_1-C_4)	10.7	8.7	12.3
Naphtha	(C_5-200°C)	15.9	9.9	22.8
Middle Oil	(200-345°C)	7.9	6.0	5.8
Heavy Oil	(345-525°C)	23.9	21.0	15.1
Redidual Oil	($>$ 525°C)	17.0	27.5	11.6
		75.4	73.1	67.6
Unreacted Coal and Ash		14.3	15.8	17.3
H_2S and NH_3		4.8	4.2	1.0
H_2O and CO_2		10.0	10.1	19.8
		104.5	103.2	105.7
Hydrogen Consumption		4.5	3.2	5.7
		100.0	100.0	100.0

Source: R.M. Eccles and G.R. Devaux, "Current Status of H-Coal Commercialization", paper presented at 2nd World Congress of Chemical Engineering held in Montreal, Canada, October 4-9, 1981. Proceedings Volume II, pp. 514-520.

Table 2-4 Operating Conditions for a Typical Day (27/3/81-28/3/81) for the H-Coal Pilot Plant (Bituminous Illinois No. 6 Coal)

	Actual	Target
Reactor Temperature, °C	452	455
Pressure, MPa	20.7	20.7
Coal Feed Rate, t/d	201	190
Space Velocity, kg dry coal/h/m^3	506	500
Slurry Oil-to-Solid Ratio	1.65	1.75
Hydroclone Recycle, % of Slurry Oil	47	60

Table 2-5 Product Yields* for a Typical Day (27/3/81-28/3/81) for the H-Coal Pilot Plant (Bituminous Illinois No. 6 Coal)

	Actual		Target	
	wt%	bbl/t	wt%	bbl/t
Hydrocarbon Gases (C_1-C_3)	10.96	-	10.68	-
Naphtha (C_4-200°C)**	22.71	1.51	18.74	1.27
Distillate (200-525°C)	23.89	1.32	28.33	1.48
Residuum Oil ($>$ 525°C)	21.62	0.85	19.00	0.78
Unconverted Coal	3.47	-	5.78	-
Ash	11.22	-	11.51	-
TOTAL		3.68		3.53

* MF basis

** In the data, HRI has included butane (C_4) in the naphtha fraction.

Source: R.M. Eccles and G.R. Devaux, "Current Status of H-Coal Commercialization", paper presented at 2nd World Congress of Chemical Engineering held in Montreal, Canada, October 4-9, 1981. Proceeding Volume II, pp 514-520.

In September 1981, the pilot plant processed Illinois No.6 coal under desired operating conditions. The run was in solvent balance and the catalyst consumption was close to the predicted value of 0.5 kg/t MF coal (or 0.6 kg/t MAF coal). In early 1982, a short experimental run was made with subbituminous coal to evaluate a new catalyst. Since then the pilot plant is processing bituminous Illinois No. 6 coal to substantiate the data for the proposed Breckindridge commercial project.

2.1.2 **Technical Review**

A technical review of the H-Coal process presented below is based on available information related to processing of lignites in the open literature and material directly received from the process developer.

2.1.2.1 Types of Coal Investigated

Different types and ranks of coal (brown, lignite, subbituminous and bituminous) have been investigated in the H-Coal process at the HRI facilities. Table 2-6 presents the breakdown of accumulated number of hours spent on bench-scale and PDU experiments with various types of coals.

From the coals tested so far, bituminous Illinois coal (27,600 hours of operation) and subbituminous Wyodak coal (12,672 hours of operation) have been investigated most extensively in the bench-scale units and PDU, and were also selected along with bituminous Kentucky No.11 coal for the H-Coal pilot-plant

Table 2-6 Accumulated Hours of H-Coal Bench-Scale and Process-Development-Unit Runs

Type of Coal	Origin/ Place	Name/ Mine	No. of Hours	Total Hours
Bituminous	Arizona	Black Mesa	597	
	Colorado	Colorado	1,959	
	W. Germany	Ruhr Area	811	
	Illinois	River King Mine	10,819	
		Misc. Seams	6,699	
		Monterey Mine	1,386	
		Orient Mine No.5	448	
		Burning Star Mine	8,247	
	Kentucky	Run-of-Mine	852	
		Kentucky No. 11	713	
	Pittsburg	Pittsburg Seam	8,452	
	U.K.	Annesley Double Coal	343	
	Utah	Utah D-Seam Coal	410	
		Kaiparowits Plateau	673	42,409
Subbituminous	Wyoming	Wyodak	12,672	
		Bighorn	1,232	13,904
Lignite	N. Dakota	Dickinson Area	551	
	Texas	Texas Lignite	705	1,256
Brown	Australia	Gelliondale Field	760	760
		Muswellbrook*	-	-

* Tested in microautoclave unit only

Source: "H-Coal Run Summary in Bench-Scale and Process Development Units", HRI R&D, October 24, 1979 operation at Catlettsburg. A recent study (EPRI Report AF-1297, Dec. 1979) conducted for the engineering and economic evaluation of a large-scale H-Coal commercial plant was also based on the utilization of Illinois No. 6 and Wyodak coals. At present, a detailed design and engineering study for a commercial plant to be located in Breckindridge County, Kentucky, using bituminous Illinois No.6 coal is being carried out.

On the other hand, operational experience with the processing of lignites has been limited. North Dakota lignite was tested (550 hours of operation) in a bench-scale unit over 15 years ago when the program was under the sponsorship of the Office of Coal Research (now U.S. Department of Energy). Texas lignite was also tested for a short period (705 hours of operation) for a private client in 1967. Since then, the process has undergone several changes and improvements, and the results of these tests may not necessarily represent the present state of development of the process.

Amongst the brown coals, laboratory experiments were made with Australian coal from the Gelliondale field in 1970. Recently, another Australian brown coal from Muswellbrook, New South Wales, was tested in a microautoclave to determine its liquefaction behaviour in the H-Coal process. On the basis of these preliminary micro-autoclave results and by comparison with other similar coals tested in bench-scale units and PDU, projections were made for a conceptual commercial plant using a computer program termed "H-Coal Commercial Plant Planning Model".

Typical analyses of some of the low-rank coals such as North Dakota lignite and Australian brown from Muswellbrook are presented in Appendix B. Analysis of other coals which have been widely tested, such as subbituminous Wyodak and bituminous Illinois coal, are also included. Comparison of these coals with Saskatchewan lignites show that none of these coals are similar to Estevan or Willow Bunch lignites. However, some of the characteristics of the Australian brown coal and subbituminous Wyodak are similar to the Saskatchewan lignites. For example, the moisture content on an as-received basis is in the range of 30-35 wt%, consequently, the extent of moisture removal and the corresponding energy requirement would be nearly the same.

With respect to ash, Muswellbrook coal has the highest amount of ash (22.6 wt% of MF coal) followed by Willow Bunch (21.0 wt% of MF coal), Estevan (15.4 wt% of MF coal) and Wyodak coal (8.6 wt% of MF coal). In terms of petrographic analysis, the Estevan lignite contains more reactive macerals (89 vol%) than Muswellbrook coal (74 vol%). Therefore, higher yields for Estevan lignite as compared to Muswellbrook coal may be expected. The upper level of inertinites in both Muswellbrook coal and Willow Bunch lignite is about the same (22 wt% to 30 wt%).

If the Wyodak coal analysis is expressed on an MAF basis, the results show that the analysis is comparable with Estevan lignite as shown below:

	Wyodak Coal	Estevan Lignite	Willow Bunch Lignite	Muswellbrook Australian Brown
Proximate Analysis, wt%				
Volatile Matter	48.8	50.0	50.1	39.5
Fixed Carbon	51.2	50.0	49.9	60.5

Ultimate Analysis, wt%				
Carbon	73.0	72.3	65.9	81.9
Hydrogen	5.5	5.3	5.1	4.8
Nitrogen	0.9	1.2	0.6	1.9
Sulfur	1.1	0.7	1.1	0.8
Oxygen	19.3	20.5	27.2	10.6

The significance of these analytical similarities and/or differences is that the liquefaction behaviour of Estevan lignite may be similar to Wyodak coal. On the other hand, the ease of operation and processibility of Estevan lignite in terms of ash handling and bottoms viscosity may be similar to Muswellbrook. For these reasons, the technical review for Estevan coal is made on the basis of the experimental results of these two coals in the H-Coal process. In some cases, data has been supplemented with that of Illinois No.6 coal for comparison purposes only. Pertinent references and comments regarding the processibility of Willow Bunch lignite are also presented.

2.1.2.2 Overall Process Configuration of a Conceptual, Commercial Plant for the H-Coal Process

A simplified flow diagram of the H-Coal process in the "syncrude" mode is presented in Figure 2-4. Some of the processing units and supporting systems (not shown) comprising an integrated commercial plant which have the similar features for any "direct" coal liquefaction process have been discussed earlier (see Figure 2-1, "Typical Block Diagram of an Integrated Plant").

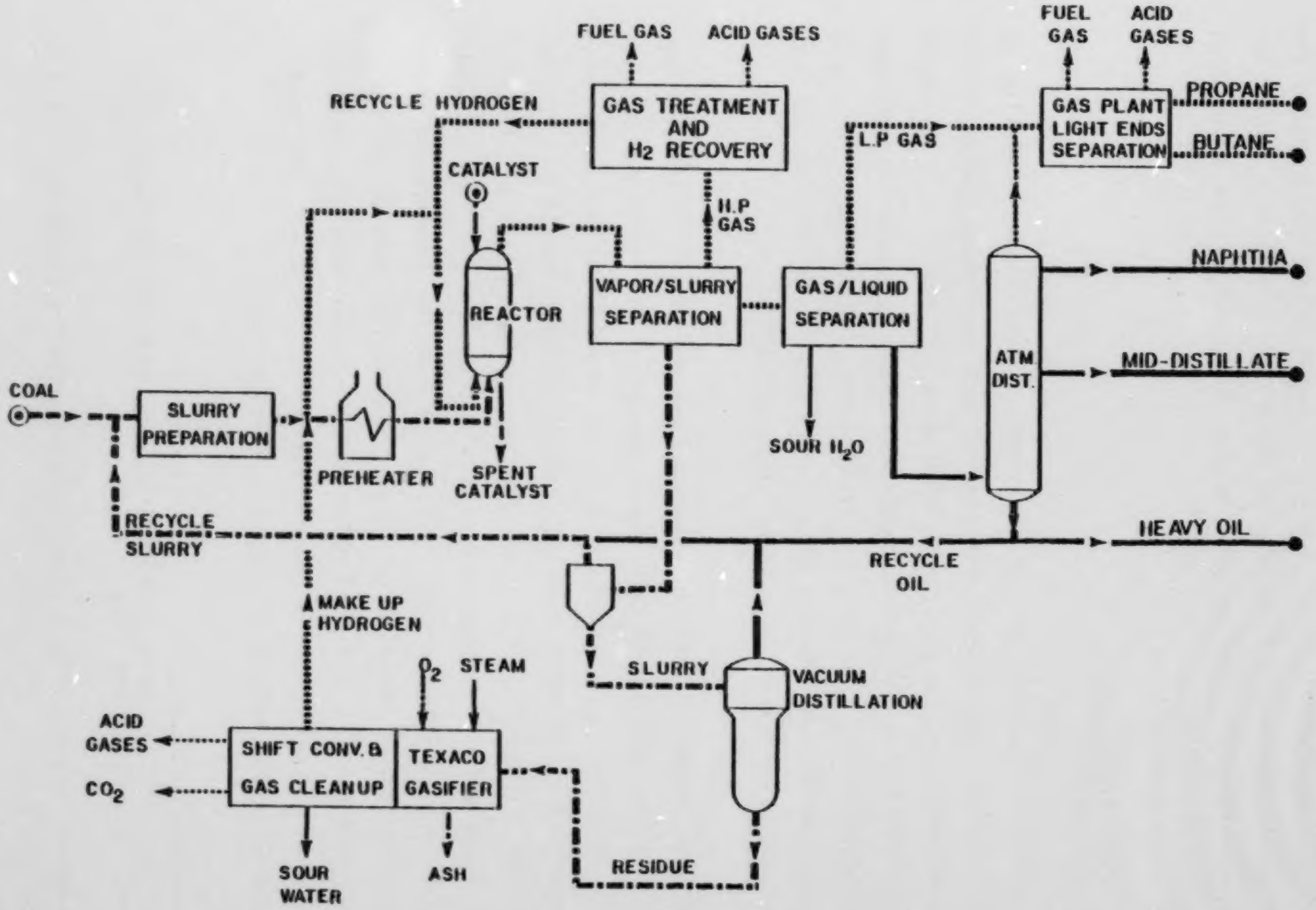

Figure 2-4 Simplified Flow Diagram of the H-Coal Process

A brief description of the main processing areas highlighting the peculiarities and differences of an H-Coal conceptual plant for liquefaction of subbituminous Wyodak and Muswellbrook brown coal are presented below.

- **Coal Preparation**

Commercial liquefaction of as-received Australian Muswellbrook coal, using the H-Coal process is not economic, unless this run-of-mine coal is pretreated to reduce its ash content to about 10 wt% MF basis. This conclusion was drawn by the H-Coal process developer, HRI, after a preliminary evaluation* of the Muswellbrook coal having high ash (22.6 wt% MF basis) and inertinite contents (19 vol% or about 30 wt%).

The high amount of ash and inertinites end up as solid residue and require correspondingly large amount of coal-derived liquids taken from the liquid product in order to make the vacuum-bottoms pumpable (as no more than 50 wt% of solids should be in the residual slurry**). Also, the high moisture content of the coal (30.0 wt% on as-received basis) would increase the cost of coal drying before liquefaction. Therefore, in the conceptual plant configuration of Muswellbrook coal liquefaction made by HRI using H-Coal Commercial Plant Planning Model it was assumed that the run-of-mine coal would be deep-cleaned to 10 wt% ash (MF basis) by benefication before it enters the plant.

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* H-Coal Evaluation for the Energy Authority of New South Wales, HRI Report, August 1981.

** Further details are discussed later, see Sections 2.1.2.3 and 2.1.2.4.

The deep-cleaned Muswellbrook coal is reclaimed from live storage and conveyed to coal crushing and drying units where it is pulverized to less than 0.5 mm and dried to about 4 wt% moisture.

For Wyodak coal, the coal preparation is nearly the same except that no deep-cleaning for removal of ash and inertinite content is required. Wyodak coal is crushed to minus 0.8-mm size and dried to about 2 wt% moisture.

- **Coal Liquefaction and Product Recovery Section**

Dried and ground coal is continuously received into the slurry preparation unit where coal is mixed with recycle oil and solid-lean slurry to form a coal/oil slurry containing about 50 wt% solids. The slurry is pumped to the reactor pressure, preheated, and fed into a catalytic ebullated-bed reactor operating at 455°C and 21 MPa. The "base" catalyst used is standard cobalt molybdate (American Cynamid 1442A). This catalyst being heavier than the reactor effluent is retained inside the reactor and is replaced once a day through an "on-line catalyst addition and withdrawal" system.

The reactor effluents are cooled in a series of heat exchangers and separated into gases, liquid and slurry products by conventional flash-separation methods. High-pressure gases are treated in an amine absorber with diglycolamine (DGA) to remove the acid gases. Treated gas is cryogenically separated into a hydrogen-rich stream which is recycled to the reactor and the remaining hydrocarbon-containing gases which are used as fuel gas for the process. Low-pressure sour gases are sent for further processing in the gas treatment section.

The condensed liquid hydrocarbon stream, collected from various flash separation vessels, is preheated and then fractionated to produce naphtha (C_5-200°C), mid-distillate (200-345°C) and heavy oil (345-525°C). Naphtha from the fractionator is stabilized (not shown) and sent to storage. Mid-distillate is pumped to storage after steam stripping and air cooling (not shown). A portion of the heavy oil is recycled to the slurry preparation unit and the remainder is sent to storage.

The slurry containing some heavy oil, unreacted coal and ash from the high-pressure flash separation, is passed through two-step pressure letdown valves to reduce the pressure from 21 MPa to about 0.7 MPa. The slurry at low pressure is then fed into hydroclones where it is mechanically divided into solid-lean (overflow) and solid-rich (underflow) streams. The solid-lean slurry is combined with recycle oil and returned to the slurry preparation unit. The solid-rich slurry is fed into a vacuum distillation tower to recover additional light oil. The residual slurry (vacuum bottoms) is gasified for the production of make-up hydrogen for the process.

- **Gas Plant and Light-Ends Recovery**

Different schemes are employed for subbituminous Wyodak and Muswellbrook brown coal. For the Wyodak case, the low-pressure gases are treated in a diglycolamine (DGA) plant for removal of acid gases. A portion of the treated gas is used as fuel gas for the process and the remaining gas is compressed to 2.6 MPa and sent to steam reforming plant (not shown) in order to produce additional make-up hydrogen for the process. No propane or butane is produced in Wyodak case because higher hydrogen consumption coupled with an insufficient amount of residual-oil yield makes it necessary to recycle all the available gases to the hydrogen production section to produce the required amount of make-up hydrogen for the process.

In the Muswellbrook case, the same processing scheme for gas treatment in an amine plant is followed. The treated gas is then sent to depropanizer and debutanizer units to produce propane and butane (LPG) as coproducts of the process.

- **Hydrogen Production**

The residual slurry from vacuum distillation is gasified in a Texaco partial oxidation unit. The gasification is carried out at elevated temperature and pressure (up to 1400°C and 4 MPa respectively) in the presence of steam and oxygen.

The product gas undergoes a shift conversion step and the shifted gas is sent to a Selexol plant where H_2S and CO_2 are selectively remove. The resulting purified hydrogen-rich stream is compressed and recycled to the coal liquefaction reactor.

- **Emission and Effluent Control Systems**

Acid-gas streams from various processing sections of the plant are combined and treated in a conventional Claus-type plant to recover elemental sulfur. The tail gases from the Claus plant are further treated using a Beavon process to recover additional sulfur and to produce an environmentally acceptable gas stream for discharge to the atmosphere.

Sour water produced from various processing sections is treated using the Phosam-W process to recover ammonia and to separate H_2S stream suitable for feeding to the Claus sulfur plant. The water stripped from the Phosam-W process is further processed to recover phenols using the dephenolization process developed by Jones and Laughlin Steel Company. Treated water is reused.

- **Utilities**

In the Wyodak case, electric power* in sufficient amount (180 MW) is produced by combined-cycle gas turbines to satisfy the in-plant requirements. These turbines are fed by residual fuel gas produced in the process.

In the Muswellbrook case, high-pressure steam and 108 MW of electricity is generated in vacuum-bottoms fired boilers. Flue gas desulfurization and electrostatic precipitators are included to control SO_2 and particulate emissions. Additional electric power (92 MW) is generated on site via process-derived fuel gas turbines and steam turbines.

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* For 25,000 t/d of as-received coal.

2.1.2.3 Product Yields and Product Efficiencies

Product yields and distributions of a self-sufficient, conceptual, commercial H-Coal plant in the syncrude mode of operation are presented in Table 2-7 for different rank of coals*.

The results show that the net liquid-product (C_5-525°C) yield decreases from 2.6 bbl/t for bituminous Illinois No.6 coal to 1.4 bbl/t of Muswellbrook brown coal on as-received basis. In other words, the yield is about 30% lower for Wyodak coal and 45% lower for Muswellbrook coal as compared to bituminous Illinois No.6 coal. This is because of the high moisture and ash contents of the low- rank coal, and that the higher-rank coals generally give more product yields to liquefaction. This may be evident by observing the net liquid yields expressed in terms of moisture-free (MF) and moisture-and-ash free (MAF) basis. Compared to Illinois coal, the net liquid yield is about 10% lower for Wyodak coal, and about 20% lower for Muswellbrook coal on MAF basis.

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* For Muswellbrook brown coal, product yields and distribution are presented considering that the as-received coal (22.6 wt% ash) is deep-cleaned to 10 wt% ash, prior to liquefaction.

If the Muswellbrook coal had not been deep-cleaned to 10 wt% ash, the liquid yield would have been only 1.2 bbl/t MF coal or 1.5 bbl/t MAF coal. This is due to two reasons. Firstly, the deep-cleaned coal yields a higher amount of liquid product on MAF basis as compared to the same coal with a higher ash content. This was confirmed by HRI in the bench-scale experiments and has been described earlier in Section 2.1.1.1, "Effect of Ash Content on Product Yields". Secondly, a large amount of recoverable oil is lost in the residual slurry to transport solids (ash and unreacted coal) from the vacuum distillation tower (to make vacuum bottoms pumpable, the residual slurry solids content and viscosity should be less than 50 wt% and 50 poise respectively).

On the other hand, the amount of lighter liquid (naphtha) produced in Muswellbrook coal is about 43% compared to 30% for bituminous Ilinois No. 6 coal. The amount of hydrocarbon gases is also higher. This is due to the higher reactivity of the lower-rank coals.

The hydrogen consumption for bituminous Illinois No. 6 coal is about 5.2 wt% MAF coal basis, 6.3 wt% MAF for Muswellbrook coal, and 6.8 wt% MAF for Wyodak coal. Compared to Illinois No. 6 coal, the hydrogen consumption of Muswellbrook is high, because this coal has low hydrogen content (4.2 wt%* instead of 4.9 wt% in Illinois No. 6 coal). In case of Wyodak coal, the difference in hydrogen consumption may be related to some extent to its higher oxygen content (17.7 wt% MF coal) as compared to Illinois No. 6 coal (8.6 wt% MF coal). The excess oxygen (9.1 wt% MF

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* 4.2% hydrogen for deep-cleaned Muswellbrook coal.

Table 2-7 Product Yields and Efficiencies in the H-Coal Conceptual Plant for Different Rank Coals
(Basis: 25,000 t/d as-received coal)

	Bituminous Illinois No.6[1]	Subbituminous Wyodak[1]	Australian Brown Muswellbrook[2]
Coal (as received)			
Moisture, wt%	10.0	30.0	30.0
Ash, wt% of MF coal	10.6	8.6	22.6 (10)*
Coal Fed to Reactor			
Moisture, wt%	2.0	2.0	4.0
Ash, wt% of MF coal	10.6	8.6	10.0
Input			
MF Coal to Reactor, t/d	22,500	17,500	14,921
MF Coal to Gasifiers, t/d	-	-	-
MF Coal to Utilities and Offsites, t/d	-	-	-
Electricity, MW	-	-	-
Ouput, t/d (bbl/d)			
SNG	-	-	597
Propane	330	-	419
Butane	534	-	495
Naphtha (C_5-200°C)	2,807 (22,983)	2,708 (21,293)	2,053 (16,054)
Mid-distillate (200-345°C)	4,170 (27,862)	2,327 (16,500)	2,375 (17,369)
Heavy Oil (345-525°C)	2,576 (13,035)	1,440 (9,132)	250 (1,496)
Sulfur	767	153	109
Phenol	62	20	-
Ammonia	256	141	153
Liquid Product yield (C_5-525°C)			
bbl/t as-received coal to plant	2.6	1.9	1.4
bbl/t MF coal to plant	2.9	2.7	2.3
bbl/t MAF coal to plant	3.2	3.0	2.6
Naphtha in Total Liquid, wt%	30.0	41.0	43.0
Thermal Efficiency, %			
Overall	71.0	62.0	67.0
Liquid Products	62.0	61.0	48.0

* For Muswellbrook brown coal, product yields and distribution are presented considering that the as-received coal (22.6 wt% ash MF coal) is deep cleaned to 10 wt% ash, prior to liquefaction.

Sources: 1. EPRI Report No. AF-1143-SR, July 1979
2. HRI Report, "H-Coal Evaluation for the Energy Authority of New South Wales", August 1981.

coal) reacts with hydrogen to form water and results in an increased hydrogen consumption of 1.2 wt% of MF coal as shown below:

	Wyodak	Illinois No. 6	Difference
• Oxygen, wt% MF coal	17.7	8.6	9.1
• Water of Reaction, wt% MF coal	17.1	6.7	10.4
• Hydrogen Consumed, wt% MF coal			
- Total	6.3	4.9	1.4
- To Form Water	1.9	0.7	1.2

For the designated Saskatchewan lignites (Estevan and Willow Bunch), the overall material balance for a conceptual, self-sufficient, integrated H-Coal commercial plant is presented in Table 2-8. The design basis is 25,000 tonnes of as-received Estevan lignite (35.0 wt% moisture) supplied to the plant. The predicted product yields are based on HRI's PDU Run No. 6 (see Table 2-2) and conceptual plant data for Wyodak coal (EPRI Report No. AF-1143-SR, July 1979). These results should be considered as preliminary and do not necessarily represent an optimum overall processing scheme.

The net liquid-product (C_5-525°C) yield for Estevan lignite is about 1.5 bbl/t of as-received lignite, 2.3 bbl/t MF lignite, and 2.7 bbl/t of MAF lignite. The "overall product" efficiency has been estimated to be 62% and the "liquid product" efficiency is 59%. For Willow Bunch lignite, the product yields are lower because of higher ash content.

Table 2-8 Product Yield and Efficiencies in the H-Coal Conceptual Plant for the Saskatchewan Lignites
(Basis: 25,000 t/d of as-received coal)

	Estevan Lignite	Willow Bunch Lignite
Coal (as received)		
Moisture, wt%	35.0	35.0
Ash, wt% of MF coal	15.4	21.0
Coal Fed to Reactor		
Moisture, wt%	2.0	2.0
Ash, wt% of MF coal	15.4	21.0
Input		
MF Coal to Reactor t/d	16,250	14,428
MF Coal to Gasifiers, t/d	-	-
MF Coal to Utilities & Offsites, t/d	-	1,822
Electricity, MW	-	-
Output, t/d (bbl/d)		
SNG	-	-
Propane	50	-
Butane	82	-
Naphtha (C_5-200°C)	2,327 (18,300)	1,420 (11,164)
Mid-distillate (200-345°C)	2,000 (14,180)	1,660 (11,768)
Heavy oil (345-525°C)	865 (5,488)	286 (1,813)
Sulfur	87	142
Phenol	17	20
Ammonia	145	88
Liquid Product Yield (C_5-525°C)		
bbl/t of as-received coal to plant	1.5	0.99
bbl/t of MF coal to plant	2.3	1.5
bbl/t MAF coal to plant	2.7	1.9
Naphtha in Total Liquid, wt%	45.0	42.0
Thermal Efficiency, %		
Overall	62.0	46.8
Liquid Products	59.0	45.7

Notes:

Product yields are based on PDU Run No.6 (see Table 2-2) and conceptual plant data for Wyodak coal (EPRI Report No. AF-1143-SR, July 1979)

Sulfur, phenol, and ammonia yields have been adjusted to reflect the differences in the elemental composition of the feed coal.

Hydrogen consumption (5.7 wt% MF coal) has been estimated considering differences in the elemental composition of the coal and assuming similar product quality.

The only utility requirement from outside sources is water (360 L/s).

No electric power is required from outside as all the electricity is generated onsite. Catalyst consumption of about 0.6 kg/t of MAF coal was considered.

2.1.2.4 Suitability/Availability of Process Inputs

Low-rank coals have certain inherent characteristics and display certain properties which affect their behaviour in coal liquefaction processes. The suitability of the designated Saskatchewan lignites and availability of process inputs such as water and catalyst are discussed under the following headings.

- **Reduction of Moisture Content**

As-received Estevan and Willow Bunch lignites contain 35.0 wt% moisture which must be reduced to 2-4 wt% before the coal is fed into the reactor. Operational difficulties were encountered in the initial PDU runs with high-moisture content Wyodak coal (see Section 2.1.1.1). The problem was resolved by installing additional air-drying mills and also by incorporating a slurry-drying system.

- **Calcium Carbonate Deposits**

Calcium carbonate deposits in the liquefaction reactor appearing as wall scale and as free-flowing agglomerates have been identified and reported in some processes (for example EDS and SRC), as one of the problems related specifically to the processing of low-rank coals. Calcium in coal is present mainly in the form of salts of humic acids within the organic matrix which decomposes during liquefaction to form calcium carbonates. The extent of the calcium carbonate deposits appear to be related to the ion-exchangeable calcium content of the coals.

In Estevan lignite, the CaO content is 11.7 wt% of ash (or 1.80 wt% MF coal) and in Willow Bunch lignite 14.3 wt% ash (or 3.0 wt% of MF coal) compared to 18.8 wt% of ash (or 1.63 wt% of MF coal) in Wyodak coal.

During extensive PDU runs with Wyodak coal, no calcium-induced deposits have been observed due to scouring action of the catalyst pellets against the reactor wall. For Estevan lignite, containing more or less the same amount of CaO as Wyodak coal calcium-carbonate deposition is not expected to be a problem. For Willow Bunch lignite, containing twice the amount of CaO, calcium-carbonate deposit may be a problem.

- **Ash Content and High Vacuum-Bottoms Viscosity**

Compared to Wyodak coal (8.6 wt% ash MF coal), Estevan lignite (15.4 wt% ash MF coal) has about twice the amount of ash. In Willow Bunch lignite, the ash content is even higher (21 wt% ash MF coal). Consequently, the vacuum bottoms (residual slurry) would have a high viscosity due to the high solid content* and the low residual-oil (> 525°C) yield. This affects the operability of the vacuum distillation tower and subsequent processing units due to increased difficulty to pump the vacuum bottoms.

In the conceptual plant configuration for Estevan lignite, dilution of the residual slurry with recoverable coal-derived liquid was considered as a means of controlling the vacuum-bottoms viscosity. The net liquid (C_5-525°C) yield dropped by 0.2 bbl/t of MAF coal and the "liquid product" efficiency dropped from 61% to 59%. In the processing of Willow Bunch lignite which has a higher ash and inertinite content, the loss of liquid product is about 1 bbl/t MAF coal and problems associated with handling of the residual slurry will be even acute.

As an alternative to dilution of residual slurry, the ash and inertinite content of the Estevan and Willow Bunch lignites may be reduced by deep-cleaning and beneficiation as suggested by HRI for Muswellbrook brown coal. The technical and economic feasibility of this alternate method should be futher investigated.

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* Solids in the residual slurry include unreacted coal, ash and inertinite matter of the coal.

- **Availability of Process Inputs**

In the H-Coal process, the main process inputs are coal, water and catalyst. It has been assumed that 25,000 t/d of as-received coal is available to the plant. On this basis, the raw water requirement for Estevan coal has been estimated to be 360 L/s. For Willow Bunch lignite, the water requirement will be in the same range.

Catalyst consumption (without regeneration) in the H-Coal reactor is typically in the range of 0.30 to 1.2 kg/t MAF coal depending on the type of catalyst used, type of coal processed, severity of operating condition employed and quality of final product required. In the conceptual plant configuration of H-Coal process, standard cobalt molybdate catalyst (American Cynamid HDS-1442A) was considered, and a catalyst replacement rate of 0.5 kg/t MF coal (0.6 kg/t MAF coal) recommended by HRI was taken. This value has been established during extensive PDU runs and has also been verified in the recent pilot-plant operation. At this rate, the annual catalyst consumption will be approximately 2.6 Gg (2,600t). The supply of catalyst in such an amount is a major concern, especially the availability of cobalt.

In this respect, HRI is carrying out some work on a laboratory scale to investigate the regeneration of catalysts as part of the H-Coal and H-Oil process development program. Some of the techniques tested are:

- Regeneration of catalyst through carbon burn step.
- Segregation of spent catalyst to recover fresher catalyst followed by carbon burn step.

- Regeneration of catalysts by chemical treatment and carbon burn step.

A definitive assessment of catalyst activity after regeneration has not yet been established. It has been estimated on the basis of bench-scale experiments that 80 to 98% recovery may be achieved. The use of regenerated catalyst can significantly reduce the make-up rate of fresh catalyst. Estimates of the catalyst make-up rate reduction cannot be made until completion of the continuous PDU run which is planned in 1982. Rough indications are that 50 to 80% reduction in catalyst make-up rates may be possible.

In addition to regeneration of Co/Mo spent catalyst, HRI is also evaluating several new catalysts, one of which (Amocat 1B) contains no cobalt. The other two catalysts (American Cynamid SN 4475, Armak PA 24268) contain nickel and molybdenum on alumina. The details were discussed earlier (see Testing of Improved Catalysts, Section 2.1.1.1).

2.1.2.5 Product Quality,Utilization and Upgrading Schemes

The product quality, utilization and upgrading schemes of H-Coal syncrude obtained from processing bituminous Illinois No.6 and subbituminous Wyodak coals have been investigated extensively. Similar studies for products obtained from lignites have not yet been carried out. Preliminary pretreatment of primary distillate products to improve their stability during storage and shipment has been investigated at HRI, but most of the work related to product analysis and upgrading tests has been carried out at Chevron, Mobil, UOP, Westinghouse, Ashland, and Dow Chemicals Co.

Table 2-9 presents the analyses of coal-derived syncrudes obtained from Wyodak and Illinois No.6 coals. The results indicate that the syncrudes are not comparable because the coal feedstocks differ significantly. Of particular interest is that Wyodak syncrude is richer in hydrogen, contains fewer heteroatoms, and is of better quality than Illinois No.6 syncrude. A similar trend in product quality may be observed in the primary distillate products as shown in Table 2-10.

Inspection of H-Coal primary distillate products reveals that although the distillates are good quality coal-derived liquids, they do contain certain impurities and therefore are unsuitable for direct utilization. For example, the oxygen content (0.56 to 1.70) is very high, and the nitrogen content (0.1 to 0.63) exceeds the usual level for petroleum-derived products. Furthermore, the products do not conform with other standard specifications for gravity and viscosity. Consequently, additional upgrading steps are required to convert them into marketable products.

The H-Coal naphtha (C_5-200°C) requires hydrotreating and subsequent catalytic reforming to produce reformate suitable as blendstock for motor gasoline. The reformate yield is about 85 vol% and has a research octane number (RON) of 103. Upgrading of H-Coal naphtha also produces net hydrogen, since the hydrogen yield in reforming (3 wt%) substantially exceeds the hydrogen consumption (1.0 to 1.5 wt%) in naphtha hydrotreatment. This hydrogen (95 mol% purity) is thus available for upgrading operations with other distillate products and/or for use as make-up hydrogen in coal liquefaction.

Table 2-9 Typical Analysis and Properties of Coal-Derived Syncrudes obtained in the H-Coal Process

	Wyodak Syncrude	Illinois No.6 Syncrude
General Properties		
°API	35	26
Aniline Point, °F (°C)	32 (0)	30 (-1)
Molecular Weight	130	147
Viscosity, mm^2/s (at 38°C)	1.22	1.65
Ramsbottom Carbon, wt%	0.23	0.29
Heptane Insolubles, wt%	0.068	0.35
Ash, ppm	20	90
Bromine Number	26	42
Elemental Analysis, wt%		
Carbon	80.20	86.96
Hydrogen	12.74	11.39
Nitrogen	0.17	0.46
Sulfur	0.041	0.32
Oxygen	0.85	1.80
Hydrogen/Carbon Atomic Ratio	1.76	1.50
TBP Distillation*, °C		
IBP/5 vol%	12/69	13/80
10/30	78/127	100/167
50	179	207
70/90	220/279	246/309
95/99	316/418	345/407

* Simulated by Chromatography

Source: D.J. O'Rear, R.F. Sullivan, B.E. Stangeland, Chevron Research Company, "Catalytic Upgrading of H-Coal Syncrudes", published in "Upgrading Coal Liquids", edited by R.F. Sullivan, American Chemical Society Symposium Series No. 156, pp. 115-144, 1981.

Table 2-10 Typical Properties of H-Coal Primary Distillate Products

	Subbituminous Wyodak			Bituminous Illinois No.6		
	Naphtha	Middle Oil	Heavy Oil	Naphtha	Middle Oil	Heavy Oil
Gravity, °API	55.8	27.8	10.9	52.3	18.5	4.9
Higher Heating Value (HHV), kJ/kg	19,893	19,012	18,187	19,798	18,457	17,767
TBP Analysis, wt%						
C_4-C_5	1.02	--	--	1.37	--	--
C_5-38°C	6.86	--	--	12.98	--	--
38-93°C	23.22	--	--	26.10	—	--
93-149°C	44.84	0.29	--	35.58	0.22	--
149-205°C	24.06	19.32	0.10	23.97	19.30	0.11
205-425°C	--	80.19	91.26	--	80.39	86.31
425-525°C	--	--	8.48	--	--	13.42
Elemental Analysis, wt%						
Carbon	84.66	86.97	88.40	85.34	88.35	89.42
Hydrogen	13.97	11.39	9.41	13.79	10.41	8.56
Nitrogen	0.10	0.22	0.46	0.24	0.47	0.63
Sulfur	0.02	0.03	0.03	0.07	0.08	0.08
Oxygen	1.25	1.39	1.70	0.56	0.99	1.31

Source: EPRI Report No. AF-1297, December 1979

The H-Coal middle oil (200-345°C) requires mild hydrotreatment in order to reduce the sulfur, nitrogen and oxygen content to the desired levels, and the hydrotreated product may be used as No. 2 fuel oil. Other options involve more severe hydrotreatment to produce diesel fuel or jet fuel. Furthermore, the middle-oil may be hydrocracked to obtain additional naphtha which may be upgraded to blendstock for motor gasoline.

The heavy oil (345-525°C) requires hydrotreating at moderate severity to reduce the sulfur and nitrogen contents to the desired level suitable for use as stationary combustion fuel. However, the viscosity is somewhat high and special storage conditions are required to prevent degradation and gum formation. Alternatively, heavy oil may be catalytically hydrocracked to produce lighter products such as LPG, naphtha, and middle oil.

2.1.2.6 Areas Requiring Further Developments

Some operational problems encountered in H-Coal runs and subsequent measures taken by the process developer for improved and reliable operation are described below.

The coal/oil slurry from the slurry-mix tank is pumped by a series of centrifugal pumps followed by high-head reciprocating pumps to raise the pressure of the slurry from 2 MPa to 21 MPa required in the H-Coal reactor. In certain PDU and pilot-plant operations, occasional erratic pumping and leakage problems have been encountered.

Improper distribution of coal and oil in preheater furnace tubes, and swelling of coal during the so-called "gel-phase" result in higher viscosity, lower heat-transfer, and higher film-temperature which are believed to be the cause of coke deposition and plugging. Erosion of return bends and flame impingment are other potential problems. In the current pilot-plant program, three design alternatives of coal/oil slurry preheaters are being investigated.

For the desired performance of the H-Coal ebullated-bed reactor, it is important to regulate the temperature, pressure, composition and feed rates of the incoming coal, oil and hydrogen gas streams. The control of these variables is dependent on the performance of equipment items mentioned earlier, such as slurry-mix tanks, slurry pump, and preheater furnaces. Furthermore, it is equally important to maintain a constant catalyst activity and bed height inside the reactor. In the pilot-plant operation, installation of an external ebullation pump and modifications in reactor control instruments has significantly improved the reactor performance.

In the initial pilot-plant operation, several shutdowns had to be made due to failure of the pressure letdown valves, inspite of the fact that valves of similar design were succesfully demonstrated earlier in the PDU runs. In the first pilot-plant run, the stainless steel valve trim lasted only a few hours. The low-grade tungsten carbide trim used in the next run ruptured in one day. Replacement with a high-grade tungsten carbide trim could only increase the valve life to 5 days. During the major turn-around completed in February 1981, valves of different design manufactured by Kieley and Mueller Inc., N.Y. (termed as "SYNFLO") were installed and since then no significant problems have been encountered.

Selection of appropriate materials of construction for equipment items operating at severe conditions is also being investigated in the pilot-plant program. Over 50 corrosion test coupons have been installed in over 200 locations in the plant. A close monitoring of erosion in equipment items such as pumps, hydroclones, and pipes is also being carried out.

2.1.2.7 Status and Commercialization Prospects

The H-Coal process has been extensively tested on bench-scale (11 kg/h of coal) and PDU scale (3.2 t/d of coal). The feasibility of the process is currently being demonstrated in a pilot plant which has a capacity of processing up to 225 t/d of coal in the syncrude mode and 545 t/d of coal in the fuel-oil mode.

Continuing research and development on the H-Coal process has led to the discovery of better catalysts, to improvement in modes of operation, and has demonstrated the versatility of the ebullated-bed reactor in processing various types of coals. The current "H-Coal Development Program" consists of laboratory R & D studies, PDU operations, process engineering and economic studies, product analysis and upgrading tests, and pilot-plant operation (1980-1982). Ashland Oil Company and HRI are seeking funds from interested U.S. Companies and foreign clients to extend the pilot-plant operation till 1984. Some of the activities that are presently being carried out are:

- Process improvements and optimization study for a commercial H-Coal plant using a linear program process model (HRI).

- Correlation of ebullated bed-reactor parameters (Conoco and HRI).
- Effect on recycle solvent as a function of operating conditions (Conoco).
- Analysis and characterization of H-Coal products including storage stability testing (Mobil).
- Upgrading coal liquids for use as power generation fuels (Mobil) and for transportation fuels (UOP).
- Utilization of coal liquids in a combustion turbine engine (Westinghouse).
- Compatibility of H-Coal liquids with petroleum products (Ashland).
- Partial oxidation of coal liquefaction residue (Texaco).
- Fluid coking of coal liquefaction residues (Exxon).
- Economic evaluation of upgrading of coal-derived liquids in comparison with petroleum - derived products (UOP).

Fluid dynamics of the H-Coal reactor system, which is a critical element of scale-up, has been extensively investigated in the PDU and in cold-flow models of PDU and pilot plant. Significant improvements in the design of reactor internals have been made and transient responses of the system under upset conditions have been quantified. Moreover, to minimize the uncertainty in reactor scale-up, the development path followed by HRI for the H-Coal process is similar to that previously demonstrated and commercialized for the H-Oil process*. The reactor diameters are shown below:

	H-Oil Reactor Diameter	H-Coal Diameter
• Bench-scale	2.0 cm	2.0 cm
• PDU	22.0 cm	15 and 22 cm
• Pilot plant	1.4 m	1.5 m
• Commercial plant	4.0 m	3-4 m

The 1.5-m diameter H-Coal reactor is currently in operation at Catlettsburg pilot plant, while the commercial-scale reactors are being designed as part of the Breckinridge H-Coal commercial plant project. An H-Coal commercial plant would have several reactors operating in parallel. In terms of the individual reactor train, the commercial scale reactor would have about ten-times the throughput as the Catlettsburg pilot plant with a diameter scale-up ratio from two to three.

Work has recently been completed on Phase-Zero of the Breckinridge H-Coal liquefaction plant. The Phase-Zero is a 9-million dollars cooperative effort between U.S. Department of Energy, Ashland Oil, and Airco Energy Co. The work included process design, cost estimation, environmental assessment and economic analysis of a plant capable of processing approximately 20,000 t/d of run-of-mine Illinois No. 6 coal to produce a nominal 50,000 bbl/d of hydrocarbon liquid products. The plant would be owned and operated by a partnership of companies headed by Ashland Oil Company. In October 1981, Bechtel joined in as a partner of the H-Coal Breckinridge project and will be responsible for process design, engineering and general contracting. However, recently, Airco Energy Inc. announced its withdrawal from the H-Coal project.

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* The H-Oil process was commercialized in late 1960 and several plants are in operation today. The largest plant is in Kuwait and has a capacity of processing up to 55,000 bbl/d of heavy-oil residuum.

An application for loan guarantees and financial support has been submitted to the U.S. Synthetic Fuels Corporation. The current schedule calls for detailed engineering design (Phase I) to begin in 1982 leading to construction (Phase II) in 1983 with initial production (Phase III) in 1988.

2.2 RUHRKOHLE/VEBA OEL PROCESS

Ruhrkohle AG and Veba Oel AG are developing a direct, catalytic coal liquefaction process for the production of transportation fuels and/or petrochemical feedstocks. The RAG/Veba process, often referred to as "New German Technology" is a modified and improved version of the classical Bergius/I.G.-Farben approach.

In the RAG/Veba process, dried and pulverized coal (1 wt% moisture, minus 1-mm size) is mixed with recycle solvent (medium/heavy oil: 40/60) and the slurried coal is further ground to about 0.2-mm size. During secondary grinding, the required amount of catalyst (6.5 wt% of MAF coal) is also added to form a coal/oil catalyst slurry containing about 40 wt% solids. The catalyst used is red mud, a waste clay obtained from bauxite processing, containing 30-40 wt% Fe_2O_3 as catalytically active component. The resulting slurry is preheated to 425°C and mixed with make-up hydrogen before feeding into the reactor operating at a pressure of about 30 MPa. In the reactor, an exothermic reaction is carried out but the temperature is maintained at 475°C by injecting cold hydrogen at different levels in the reactor. The effluents are cooled and separated into gases, liquid and residual oil products by conventional flash separation and distillation techniques. The primary products are LPG, naphtha (C_5 - 200°C), mid-distillate (200° - 325°C) and heavy oil ($>$ 325°C). By-products include sulfur, ammonia and crude phenols.

In the RAG/Veba process described above, several changes and improvements have been incorporated over the classical Bergius/I.G.-Farben process, some of them are mentioned below:

- Vacuum distillation and flashing methods are used for solid/liquid separation instead of centrifuging and coking.

- Asphaltene-free distillate oil is used as recycle solvent instead of residual oil containing asphaltenes and solid residues.

- Partial oxidation of vacuum bottoms (residues) to produce make-up hydrogen as a substitute for gasification of coke purchased from outside source.

The improvements have resulted in reduction of reactor operating pressure from 70 MPa to 30 MPa; improved heat recovery and thermal efficiency; and lower hydrogen consumption* (5.5 wt% instead of 7.5 wt% MAF coal).

2.2.1 History and Development of the Process

In 1913, "direct" catalytic hydrogenation of coal was demonstrated on a laboratory scale by Professor Bergius and by 1927, a first large-scale demonstration plant was built and operated by I.G. Farben in Leuna, Saxony, Germany. By 1945, there were 12 coal liquefaction plants built in Germany having a total annual capacity of 4 million tonnes of products. Different types of feedstocks such as bituminous coal, brown coal, lignites, tar from bituminous coal processing and petroleum residues were used to produce mainly motor gasoline and aviation fuels.

Non-catalytic "direct" hydrogenation of coal was also developed in Germany as the Pott-Broche process and one plant was operated until 1945. The feedstock for this plant was bituminous coal and the main product was an extract which was used for the manufacture of electrodes for aluminium production. The annual production capacity of this plant was about 30,000 tonnes of extract.

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* Value of hydrogen consumption refers to coal liquefaction section only, i.e. without product upgrading and refining section.

2.2 RUHRKOHLE/VEBA OEL PROCESS (cont'd)

2.2.1 History and Development of the Process (cont'd)

Subsequent to World War II, gradually all the coal liquefaction plants were closed down for economic reasons and no further development work was done until the oil crisis of 1973. Following this event, four companies in the Federal Republic of Germany reactivated the work on coal liquefaction, namely Ruhrkohle AG, Veba Oel AG, Saarbergwerke AG and Rheinische Braunkohlenwerke AG.

Ruhrkohle AG, the largest coal mining company in the Federal Republic of Germany considered that the catalytic hydrogenation as developed by I.G.Farben offered flexibility for improvements and seemed most suitable for producing motor fuels, heating oil and chemical feedstocks. A bench-scale experimental unit which has a maximum capacity of processing up to 500 kg/d of coal was set up in 1976 at Bergbau-Forschung GmbH laboratories in Essen with financial support of North-Rhine Westphalia Government.

At the same time, Veba Chemie AG (now Veba Oel AG) also initiated work on coal technology development with financial assistance from the North-Rhine Westphalia Government. Initial efforts were directed towards developmment of a non-catalytic, two-stage coal liquefaction process based on Pott-Broche technology. A bench-scale experimental unit with a maximum capacity of 120 kg/d was installed at Veba Oel AG laboratories at Scholven. But later on, this process was discarded in favor of direct, catalytic coal liquefaction based on the classical Bergius/I.G.-Farben approach because the objectives of the coal liquefaction program had changed. Instead of producing mainly fuel oil, Veba Oel AG decided to produce mainly lighter fractions such as naphtha

(C_5 - 200°C) and mid-distillate (200 - 325°C) for eventual use as transportation fuels.

In 1977, Ruhrkohle AG and Veba Oel AG enterred into a cooperative agreement to jointly design, construct and operate a 200-t/d pilot plant at Bottrop, North-Rhine Westphalia, in order to advance the common technology up to a demonstration stage.

In the following sections, a summary of the work and progress made by Ruhrkohle AG and Veba Oel AG towards development of their version of the Bergius/I.G.-Farben process is presented.

2.2.1.1 Bench-Scale Experimental Program

The modified RAG/Veba process has been developed on a 500 kg/d bench-scale unit at Bergbau-Forschung GmbH facilities, Essen, and in a 120 kg/d bench-scale unit at Veba Oel laboratories, Scholven. Analysis and upgrading tests on coal-derived liquid have been carried out by Veba Oel AG.

Figure 2-5 is a schematic diagram of the bench-scale unit at Bergbau-Forchung GmbH which has the maximum capacity to process up to 500 kg/d of coal. Two reactors in series are shown, but normally only one is in operation. The second reactor can easily be added if the maximum throughput of 500 kg/d of coal is desired. Both reactors are 2 m in height and have a volume of about 11 L each. Both of them are equipped with vertically reciprocating agitation devices, which are magnetically driven to prevent settling of the particles. Internal reaction temperature is controlled at about 475°C by means of a surrounding heater to achieve isothermal temperature distribution (± 1°C).

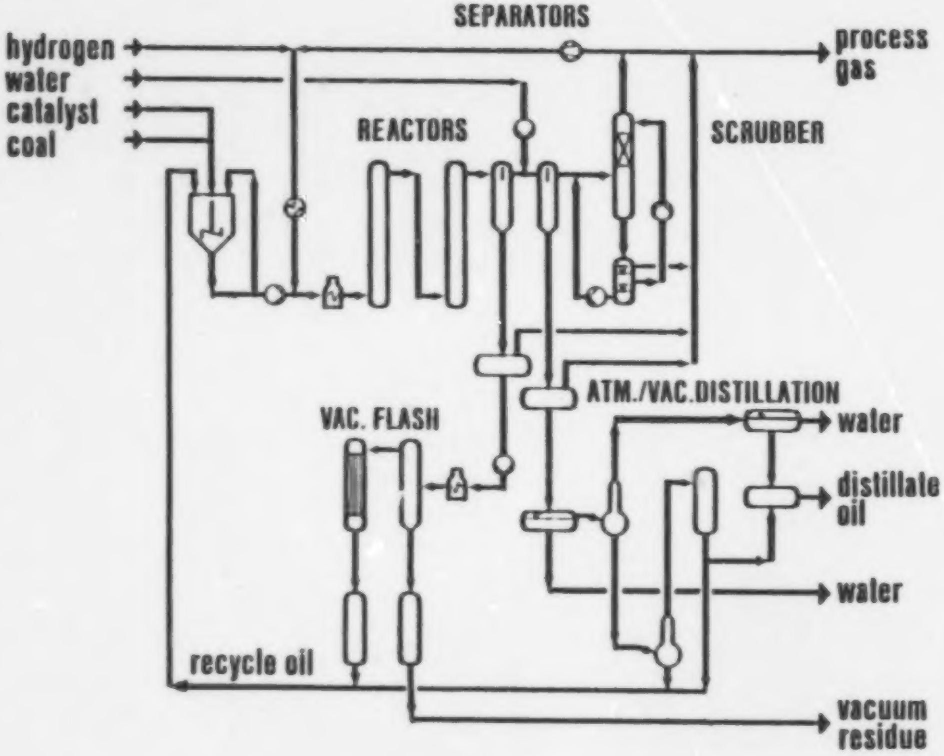

Figure 2-5 Schematic Diagram of the Bench-Scale Experimental Unit at Bergbau-Forschung GmbH (250 to 500 kg/d of coal)

Source: B. Strobel, I. Romey and G. Kolling, Bergbau-Forschung GmbH, "German Routes to Coal Hydrogenation", paper presented at Institute of Gas Technology Sponsored Symposium on Advances in Coal Utilization Technology at Louisville, Kentucky, May 14-18, 1979

Table 2-11 Typical Operating Conditions and Range of Product Yields for "Standard" German Bituminous Coal Obtained from the Bench-Scale Experimental Unit at Bergbau-Forschung GmbH

Operating Conditions	
Coal Particle Size, mm	< 0.1
Coal Throughput, $kg/h/m^3$	500 to 750
Red Mud Catalyst Consumption, kg/t MAF coal	65
Solids Concentration in Slurry, wt%	40 to 50
H_2 Partial Pressure at Preheater Inlet, MPa	25
Preheater Outlet Temperature, °C	420
Reactor Temperature, °C	475
Total Reactor Pressure, MPa	30
Make-up Hydrogen, m^3/kg MAF coal	1.0
Product Yields, wt% MAF coal	
Hydrocarbon Gases (C_1-C_4)	20.3 to 22.7
Naphtha (C_5-200°C)	13.3 to 18.4
Mid-Distillate (200-325°C)	20.3 to 34.4
Heavy Oil ($<$325°C)	0 to 4.0
Residue Slurry	
- Unreacted Coal	5.9 to 8.3
- Residual Oil ($<$325°C)	20.0 to 29.3
H_2O, NH_3, H_2S, CO_2, NaCl	9.3 to 11.4
Hydrogen Consumption	5.4 to 6.0

Source: J. Langhoff, E. Wolowski, G. Escher, and H. Hosang, "Status of German Technology on Coal Hydrogenation Development by Ruhrkohle AG and Veba Oel AG," paper presented at the Annual Fall Conference of the Canadian Society for Chemical Engineering at Sarnia, Ontario, 30 Sept. - 3 Oct., 1979.

To date, extensive bench-scale experimental runs have been carried out at Bergbau-Forschung GmbH and Veba Oel laboratories using different types and ranks of German coals and some coals from other countries. However, the main effort has been directed towards evaluation of bituminous coal from the Ruhr area. Analysis of this "standard" coal is given in Appendix B. Typical operating conditions and range of product yields obtained are summarized in Table 2-11.

2.2.1.2 Pilot Plant at Bottrop

To demonstrate the process on a larger scale, a pilot plant of 200 t/d coal capacity was constructed at Bottrop. This pilot plant was directly scaled up from the 500 kg/d bench-scale unit based on new bench-scale data and past operating experience of old I.G. Farben commercial plants. Construction of the plant was initiated in May 1979. Completion and commissioning of the plant originally scheduled for February 1981 was delayed until July 1981.

During the pilot-plant program (1981-83), bituminous coal from the Ruhr area will be processed. However, the plant has been designed to accept a wide range of coals. A simplified block diagram of the pilot is given in Figure 2-6.

The pilot plant is located near the coal mine and is adjacent to the Prosper coke-oven plant which is owned and operated by Ruhrkohle AG. Coal is received from the coking plant and the granulated residue is delivered back via a connecting belt-conveyor system. Pipelines from the coking plant also supply fuel gas, water and steam to the Bottrop plant, while the raw gases and sour water produced are returned for treatment to the coking plant. Make-up hydrogen for the process is tapped from the pipeline of Chemische Werkes Huls AG passing by the plant site.

Expected product yields of the pilot plant from "standard" bituminous coal (Ruhr area) are given in Table 2-12. The product yields shown have been predicted by Ruhrkohle AG and Veba Oel AG on the basis of bench-scale experimental data obtained by Bergbau-Forchung GmbH.

2.2.2 Technical Review

2.2.2.1 Types of Coals Investigated

Different types of feedstocks such as bituminous coal, brown coal, lignite, tar from bituminous coal processing and petroleum residues have been tested and used in the classical I.G. Farben processes. In the RAG/Veba process, bituminous coal from the Ruhr area, referred to as "standard" coal by the process developers, has been tested extensively. Some preliminary investigations have been made with other coals.

The Canadian subbituminous coal (Forestburg) was tested in autoclave unit by Veba Oel AG in 1981 on request of SNC for a separate study.

A series of preliminary tests was made for Canadian coal Forestburg and German coal (Ruhr area) under identical sets of conditons because the liquefaction behaviour of the latter is well known in both the batch autoclave and the continuous bench-scale unit. Based on the yield and analysis of the autoclave products (gases, condensate, and residues), the product

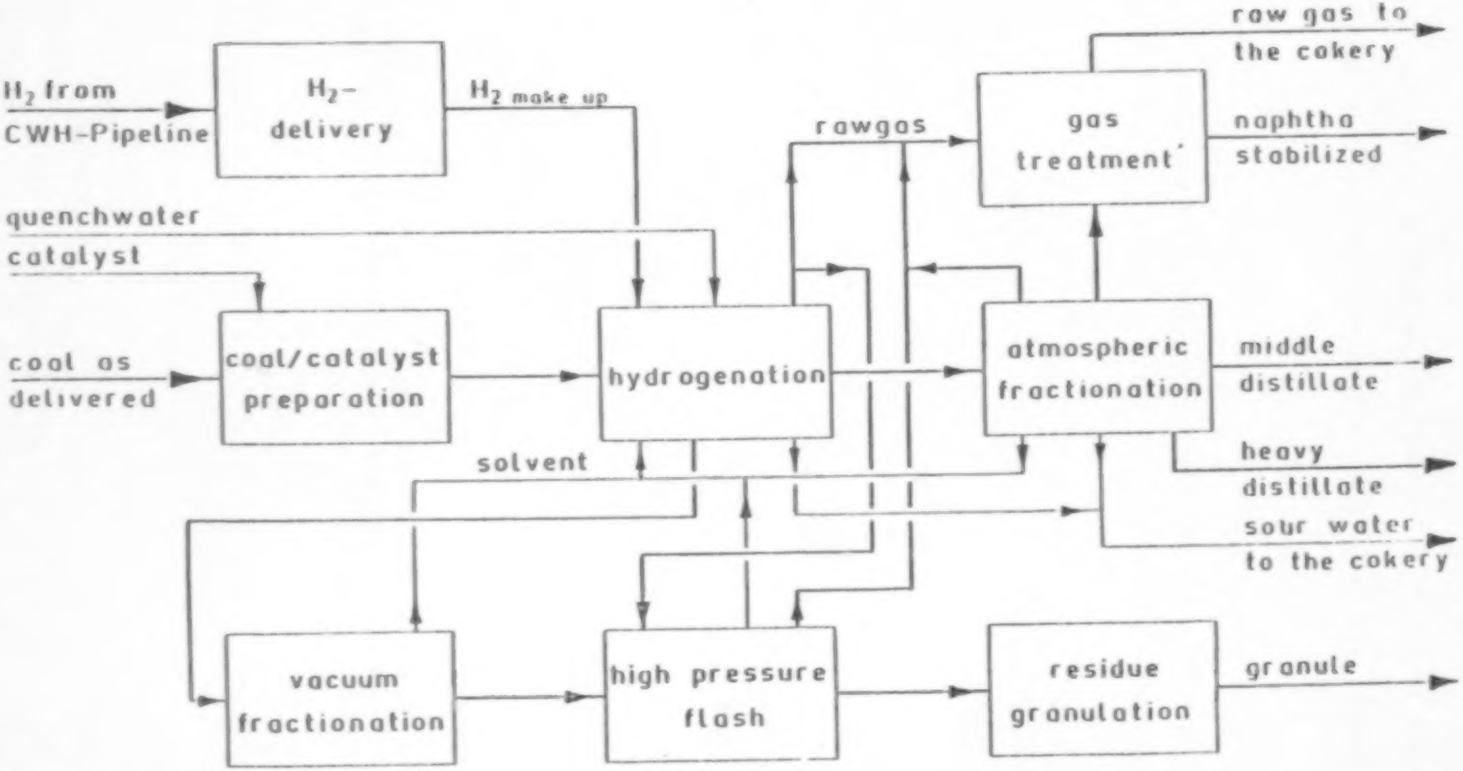

Figure 2-6 Block Diagram of the Bottrop Pilot Plant (200 t/d of coal)

Source: R. Specks et al, "The Coal Hydrogenation Plant at Bottrop", Sept. 1980

Table 2-12 Feed and Product Streams of the Bottrop Pilot Plant

(Coal: Standard German Bituminous Coal from Ruhr Area)

Feedstock/Input	
Coal (MAF basis)	200
Make-up Hydrogen, m^3/d	220,000
Raw Water, t/d	41.2
Catalyst (Fe_2O_3), t/d	4.0
Products/Output, t/d	
Gases (C_1-C_4, H_2S, CO_x, NH_3)	61.3
Naphtha	24.4
Mid-Distillate (Middle Oil)	69.0 to 74.7
Heavy Oil	-
Residues (Unreacted Coal, Catalyst, etc.)	68.6 to 74.3

Source: R. Specks, J. Langhoff, E. Wolowski, Ruhrkohle Oel und Gas GmbH, and g. Escher, Veba Oel Entwicklung - Gesselschaft GmbH, "The Coal Hydrogenation Plant at Bottrop", Sept. 1980.

Table 2-13 Product Yield and Distribution of Canadian Subbituminous Coal and Comparison with German Standard Coal.

	German Standard Coal	Canadian Subbituminous Coal
Product Yields, wt% MAF coal		
Hydrocarbon Gases		
Methane	7.4	8.0
Ethane	4.0	4.9
Propane	2.3	3.4
i-Butane	0.2	0.3
n-Butane	0.7	0.7
Pentane	0.3	0.5
Unsaturated	0.1	0.2
	15.0	18.0
Hydrocarbon Liquids		
< 200°C	21.0	10.0
200 - 350°C	38.0	34.0
250 - 500°C	13.0	15.0
> 500°C	6.0	7.0
	78.0	66.0
Water of Reaction	4.0	15.0
Hydrogen Consumption	5.2	5.0

Source: Veba Oel AG, "Autoclave Tests of Liquefaction Behaviour of Canadian Subbituminous Coal", report prepared by Veba Oel AG for SNC Inc., July 1981.

The slurry is then pumped via high-pressure reciprocating pumps to 30 MPa and mixed with recycle and make-up hydrogen streams. The coal/oil catalyst slurry and hydrogen mixture is preheated to 425°C before being fed into a series of coal hydrogenation reactors. The exothermic hydrogenation reaction occurring in the reactors which is controlled by introducing quenching hydrogen into the reactors so that the temperature does not exceed 475°C.

The effluents from the reactor are separated into gases, liquid hydrocarbons, sour water, and slurry streams by conventional separation methods.

The condensed liquid hydrocarbon stream after gas/liquid separation is preheated and then fractionated in the atmospheric distillation tower to produce primary distillate products: naphtha, mid-distillate or middle oil, and heavy oil. The naphtha is sent to stabilizer (not shown) before product upgrading. When processing low ash coals, a portion of the heavy oil stream is recycled to the coal/oil slurry preparation unit and the remainder may be sent for product upgrading. For high ash Saskatchewan lignites, all the heavy oil is recycled in order to adjust the oil-to-solid ratio in the residual slurry.

The solids containing slurry is fed into the vacuum distillation tower to recover a portion of distillate oil which is recycled to the coal-oil slurry preparation unit. The vacuum tower bottoms (residual slurry) containing about 50 wt% solids (unreacted coal, ash and catalyst) are pumped to the hydrogen production section.

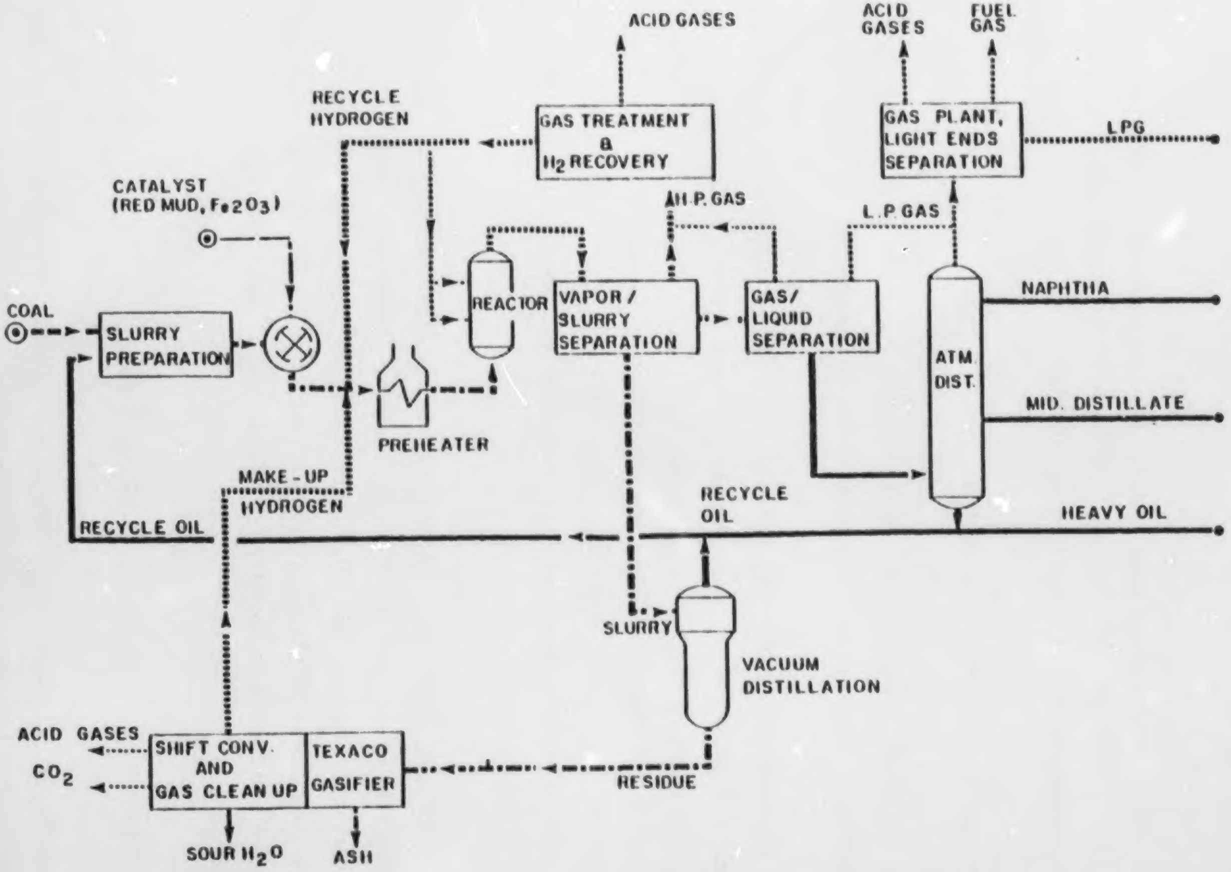

Figure 2-7 Simplified Flow Diagram of the RAG/Veba Process (New German Technology)

The residues are gasified in Ruhrkohle/Ruhrchemie/Texaco gasifier.

Sour water containing crude-mixed phenols and dissolved ammonium salts is sent to the effluent control unit. Acid gases are sent to a Claus plant for recovery of sulfur.

2.2.2.3 Product Yields and Efficiencies

The product yield and distribution of a self-sufficient, conceptual, commercial RAG/Veba plant is presented in Table 2-14.

The results show that the net liquid product (C_5-525°C) yield is about 2.3 bbl/t of as-received bituminous German coal compared to 1.0 bbl/t of as-received Estevan lignite. The yield is almost 40% less for Estevan lignite and this is because of its high-moisture and ash content. On MAF basis, the yield for bituminous German coal and Estevan lignite are 2.5 bbl/t and 1.7 bbl/t, respectively.

The "overall product" efficiency for bituminous German coal is 60.5% compared to 55.7% for Estevan lignite. The "liquid product" efficiency are 49.1% and 44.0% respectively. For Willow Bunch lignite with higher ash content, even further reduction in product yield and product efficiencies may be expected in comparison to bituminous German coal.

2.2.2.4 Suitability/Availability of Process Inputs

Saskatchewan lignites have certain inherent characteristics and properties which may affect their behaviour and processibility in RAG/Veba process as discussed below:

Table 2-14 Product Yield and Efficiencies in the RAG/Veba Conceptual Plant (Basis: 25,000 t/d "as-received" coal)

	Bituminous German Coal	Estevan Lignite
Coal (as received)		
Moisture, wt%	4.3	35.0
Ash, wt% of MF coal	4.5	15.4
Coal Fed to Reactor		
Moisture, wt%	1.0	2.0
Ash, wt% of MF coal	4.5	15.4
Input		
MF Coal to Reactor, t/d	21,165	12,505
MF Coal to Gasifiers, t/d	-	-
MF Coal to Utilities, t/d	2,759	3,745
Electricity, MW	230	160
Output, t/d (bbl/d)		
SNG	-	-
Propane/Butane (LPG)	1,818	952
Naphtha (C_5-200°C)	2,929 (20,960)	1,058 (7,570)
Mid-Distilate (200-325°C)	6,972 (44,290)	3,553 (22,570)
Heavy Oil (325-525°C)	201 (1,200)	-
Sulfur	81	57
Phenol	-	-
Ammonia	263	116
Liquid Product Yield* (C_5-525°C)		
bbl/t as-received coal to plant	2.3	1.0
bbl/t MF coal to plant	2.4	1.5
bbl/t MAF coal to plant	2.5	1.7
Naphtha in Total Liquid, wt%	29.0	23.0
Thermal Efficiency, %		
Overall	60.5	55.7
Liquid Products	49.1	44.0

Notes: For bituminous German coal, product yield and efficiencies are calculated using conceptual plant data (19100 short ton/day of coal) and information package supplied by Veba Oel AG, Ref. No. 5S3/hs-ku, 27 July 1981. For Estevan lignite, product yield and efficiencies are based on the assumption that its liquefaction behaviour is similar to Forestburg coal on MAF basis (see Table 2-13). Sulfur, phenol, and ammonia yield have been adjusted to reflect the differences in the elemental composition of feed coals.

* Liquid product yield for both bituminous German coal and Estevan lignite have been adjusted to zero electricity import using 10,000 Btu/kWh (10,500 kJ/kWh) and fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

- **Reduction of Moisture Content**

The moisture content of as-received Estevan and Willow Bunch lignites (35 wt%) must be reduced to less than 2 wt% before feeding the coal in the reactor. In the RAG/Veba process, the total operating pressure employed is already very high, 30 MPa. Failure to reduce the moisture to the desired level (< 2 wt%) will further increase the severity of the operating conditions. In addition, presence of excess moisture may create oil/water emulsions and foaming problems.

- **Ash Content and High Vacuum-Bottoms Viscosity**

Compared to bituminous German coal (4.52 wt% ash), Estevan lignite (15.4 wt% ash) has about 3.5 times and Willow Bunch lignite (21 wt% ash) 4.6 times the amount of ash. Moreover, in the RAG/Veba process, red mud is used as a catalyst in the amount of 6.5 wt% MAF coal. Consequently, the vacuum bottoms (residues) have a high viscosity due to the high solid content (ash, unreacted coal and catalyst) and low residual oil (> 325°C) yield.

In the conceptual plant configuration for Estevan lignite, dilution of residual slury with recoverable heavy oil (325-525°C) was considered as a mean of controlling the vacuum-bottoms viscosity.

- **Availability of Process Inputs**

In the RAG/Veba process, the main process inputs are coal, water, catalyst and electricity. It was assumed that 25,000 t/d of as-received coal (Estevan or Willow Bunch lignites) is available to the plant.

On this basis, the water requirement for Estevan or Willow Bunch lignite is about 400 L/s. A catalyst consumption rate of 6.5 wt% of MAF coal feed to reactor amounts to 213,200 tonnes per year. The electricity requirement is 160 MW.

2.2.2.5 Product Quality, Utilization and Upgrading Schemes

An on-going research program directed towards evaluation and upgrading of coal-derived liquid products has been initiated by Ruhrkohle AG and Veba Oel AG as a part of their coal liquefaction program. The syncrude used in these experiments was supplied by Bergbau-Forschung GmbH from the bench-scale coal liquefaction experiments with bituminous German coal from the Ruhr area. When the Bottrop pilot plant (200 t/d of coal) is in operation, the primary distillate products from the plant will be tested by Veba Oel AG on a semi-commercial scale in their laboratories in Scholven.

Table 2-15 presents a typical analysis of primary distillate products obtained from processing of bituminous German coal.

Upgrading of naphtha to motor gasoline and BTX with phenols and other base compound as byproducts has been investigated by Veba Oel AG. Some experiments have been carried out in a two-stage mixer-settler to extract phenolic compounds present in naphtha (up to 20 wt%) before naphtha hydrotreatment. A sodium hydroxide solution (12 wt% NaOH) at 48°C with a solution-to-naphtha ratio of 1/4 was used to achieve up to 90% extraction.

Naphtha hydrotreating tests have been made in a two-stage, bench-scale catalytic 1-liter reactor. In the first stage, cobalt molybdate was used as a catalyst and in the second stage nickel molybdate was employed. Temperatures of 360°C and 400°C in the first and second stages, respectively, and corresponding pressure of 6 MPa and 10 MPa were maintained. A space velocity of 2 kg/h/kg (catalyst) in each stage is necessary to achieve the required reformer feed specification, i.e. reduction of sulfur and nitrogen contents to values less than 1 ppm. Reforming of straight-run hydrotreated naphtha was done at a temperature of 490°C, a pressure of 2.5 MPa and a space velocity of 1.5 kg/h/kg (catalyst) using a platinum catalyst. The reformate is highly aromatic and is a good blendstock for motor gasoline having research octane number of 99.

Mid-distillate has been tested for the production of diesel fuel and heating oil. The sulfur content of the mid-distillate stream is 80 ppm, which is within the target specification of sulfur (3000 ppm) for heating oil and diesel fuel in Germany. However, the nitrogen and oxygen contents are very high, 0.85 wt% and 2.6 wt% respectively, and the cetane number is almost zero.

For heating purposes, the untreated mid-distillate fraction can be used as blendstock for petroleum-derived heating oil. Burner tests were made by Vea Oel AG and the results indicated that up

Table 2-15 Analytical Data of Distillate Products obtained from the Bench-Scale Experimental Unit

	Naphtha (Light Oil)	Mid-Distillate	Heavy Oil
Specific Gravity at 15°C	0.880	0.992	1.060
Gravity, °API	29.2	11.2	1.99
Boiling Range, °C			
Initial Boiling Point	76	210	311
10 Vol%	115	223	343
50 Vol%	164	249	371
End Point	212	319	N.A.
Elemental Analysis, wt%			
Carbon	85.05	88.15	88.15
Hydrogen	10.80	8.80	7.30
Nitrogen	0.24	0.85	1.45
Sulfur	0.03	0.08	0.39
Oxygen	3.60	2.60	2.2
PONA Analysis, wt%			
Paraffins	38.3	18.7	7.9
Olefins	-	-	-
Naphtenes	13.0	17.0	20.0
Aromatics	48.7	64.3	72.1

Source: G. Escher et al., "Ruhrkohle/Veba Oel Technology for Hydrogenation of Hard Coal", paper presented at Annual Fall conference of Canadian Society of Chemical Engineering, Edmonton, Alberta, Oct. 19-22, 1980.

to 10% of untreated coal-derived oil can be directly blended, but some chemicals must be added to the mixture to prevent degradation of the product during storage.

On the other hand, if the mid-distillate fraction is used without dilution, then mild hydrotreatment (390°C, 10 MPa, 1.2 kg/h/kg) is required to remove the heteroatoms and stabilize the product. For diesel fuel purposes, severe hydrotreatment (420°C, 28 MPa, 1.0 kg/h/kg) must be employed because the aromatic rings must also be saturated and partially cracked in order to increase the cetane number. The catalysts used were nickel molybdate and nickel tungsten respectively.

The final products obtained from low-severity hydrotreatment of mid-distillate have low nitrogen (< 1000 ppm) and sulfur (< 16 ppm) contents, but the aromaticity (39 wt%) and the specific gravity (0.92) are still high. Therefore, existing burner designs will require modification if this oil is to be used as heating fuel.

As a result of high-severity hydrotreatment, the aromatic compounds were completely hydrogenated, but very little cracking of the naphthenic rings occurred because the cetane number of 30 is still low. Therefore, Veba Oel AG is planning to carry out further investigations using other catalysts with higher cracking activities.

Upgrading tests on heavy oil (325-525°C) have not yet been carried out. The heavy oil is highly aromatic (72 wt% aromatics) and contains high amounts of nitrogen (1.45 wt%), oxygen (2.2 wt%) and sulfur (0.39 wt%). Besides for some special applications, the heavy oil would be mainly hydrocracked to produce light oil and mid-distillate. These hydrocracked products will be further upgraded to gasoline, diesel fuel and heating oil as described above.

2.2.2.6 Areas requiring Further Development

The areas of technical concern and some operational problems encountered in RAG/Veba process are described below. Steps taken by process developers and level of success achieved is also discussed.

The coal/oil/catalyst slurry from the slurry preparation tank is pumped to a pressure of about 30 MPa by a series of reciprocating pumps. Because of the abrasive nature of the slurry and the high-pressure pumping, special wear-resistant pumps are required In the Bottrop pilot plant, three slurry-feed pumps have been installed, two for operation and one as spare. For test purposes two of these pumps have the pistons operating vertically and third pump has the pistons operating horizontally. In the remote valve box of the slurry-feed pumps, different designs and ma-materials of construction have been used for the inlet and outlet valves installed to investigate the effects of corrosion and erosion on their performance.

The high-pressure slurry from the reciprocating pumps is mixed with make-up hydrogen and the resulting mixture is preheated to about 420°C before it is fed into the reactor. When the slurry is heated, the coal particles swell considerably. During this so called "gel phase", the viscosity of the slurry increases rapidly and reaches a maximum value between 250 and 300°C and then decreases due to partial decomposition of the coal. During the high-viscosity range, coke deposition inside the tubes may occur resulting in lower heat-transfer rates.

To partially compensate for the low heat-transfer rates, hydrogen is premixed with slurry (before preheating) to improve the heat transfer and fluid flow properties of the slurry. Two design alternatives of slurry preheaters will be evaluated during the pilot plant operation at Bottrop. The first type of preheater referred to as "convection heater" is a heat exchanger in which the slurry flowing through the tubes is heated by circulating hot air in the shell. The rate of heat transfer in this heater is low which is expected to prevent the slurry from coking in the tubes. The disadvantages are its high cost and size. The second type of the preheater is a "gas-fired furnace" similar to those used in oil refineries. Due to higher temperatures, the risk of coke deposition is also higher but the advantages are the lower capital cost and smaller size of the preheater. One of the objectives of the Bottrop pilot plant operation is to define the operating conditions in order to maximize the heat-transfer rates and minimize the coke deposition.

The coal liquefaction reactor used in the RAG/Veba process is an upflow, tubular reactor. In the initial bench-scale experiments it was found that for the required residence time, the velocity of the reactor content was not sufficient to carry all the coal particles up and out of the reactor. To avoid settling of the heavier coal particles, a magnetically driven vertically reciprocating agitation device was installed. In the Bottrop pilot plant, three reactors in series are provided instead of constructing a single large vessel. This provides sufficient residence time and at the same time allows higher reactor velocity to prevent settling of the heavier particles.

Another aspect which is equally important is the control of the temperature inside the reactor. Coal hydrogenation is exothermic and highly sensitive to temperature variations. An increase of 10°C would double the rate of reaction and may lead to temperature runaway problems. In the bench-scale experiments, the reactor was electrically heated and internal temperatures were controlled to achieve isothermal temperature distribution of less than 1°C. In the Bottrop pilot plant, the temperature will be controlled by injecting cold hydrogen at different levels in the reactor. The performance of the reactor is yet to be ascertained.

Heat exchangers are installed for cooling the reactor effluents and to recover the process heat. The main concern in this area is erosion and corrosion, especially in the heat exchangers operating at high temperature and pressure. During the bench-scale experiments, deposition of ammonium salts occurred in the heat exchanger. To prevent this, water was injected into the stream before cooling to keep the salts dissolved. In the Bottrop pilot plant, the performance of this equipment item is still subject to demonstration.

The slurry from the hot separator contains 65 to 75 wt% oil which in turn contains about 15 to 25 wt% volatile fractions. To recover the maximum amount of volatile fractions, five alternate arrangements of flash separation followed by vacuum distillation have been incorporated for testing in the Bottrop pilot plant. In three of these alternatives, the number of flash separation steps may be varied. In the remaining two arrangements, the slurry bottoms from the hot separator are stripped with hydrogen before secondary flash separation.

In all cases, the slurry is fed into the vacuum distillation tower to recover recycle oil as overhead product and to yield a solid-rich slurry containing no more than 50 wt% solids. Handling of the highly viscous, solids-containing abrasive slurry is another area of concern.

2.2.2.7 Status and Commercialization Prospects

Ruhrkohle AG and Veba Oel AG have been engaged in the development of their modified version of the Bergius/I.G.-Farben process since 1974. During this period of time, the process has been demonstrated on a bench-scale unit (250 to 500 kg/d of coal) located at the Bergbau-forschung GmbH facilities at Essen, and a 120 kg/d unit at the laboratories of Veba Oel AG at Scholven. Extensive bench-scale tests have been conducted only with bituminous German coal ("standard" coal) from the Ruhr area.

The primary coal-derived distillate products obtained in bench-scale experiments by Bergbau-Forschung GmbH, have been analyzed and tested by Veba Oel AG. The main target of the upgrading work was the production of motor gasoline, diesel fuel, jet fuel and heating oil. In addition, extraction and separation of petrochemical products such as BTX, phenols and polynuclear compounds have also been examined. So far, upgrading tests have been conducted in small-scale apparatus using naphtha and mid-distillate (middle oil) fractions. Tests on heavy oil has not yet been carried out.

On the basis of the bench-scale results and the past operating experience with the old I.G. Farben plants in Germany, Ruhrkohle AG and Veba Oel AG are operating a 200 t/d pilot plant at

Bottrop. This plant was directly scaled up from the 1/2-t/d bench-scale unit and is in operation since March 1982. A three-year operation is planned and one of the major objectives of the pilot plant program is to demonstrate the long-duration and trouble-free operation of the process.

The total cost of the pilot plant and three years projected operation (1981-83) is estimated to be DM 400 million. A major portion of this cost will be financed by the North-Rhine Westphalia Government. The remaining portion of the cost will be provided by Ruhrkohle AG and Veba Oel AG in the ratio of the 60:40.

The final objective of the process development program is to advance the RAG/Veba coal liquefaction technology towards commercialization in the 1990's. Two independent proposals have been submitted by Ruhrkohle AG and Veba Oel AG to the Federal Ministry of Research and Technology in Bonn, namely:

- Ruhrkohle AG: Commercial plant (6 million tonnes per year of coal) for the production of 2 million tonnes of primary liquid products from bituminous coal at a site in the Ruhr area. The by-products are 0.6 million tonnes of LPG and about 1.8 billion cubic metres of SNG. The total estimated investment would be about DM 4.0 billion.

- Veba Oel AG: Commercial plant processing about 6 million tonnes per year of coal, or oil residues, or a mixture of coal and oil residues to produce about 2 million tonnes per year of primary liquid products. The location of this plant would be either in the Ruhr area or on the North Sea Coast of

Germany. The investment would be the same, about DM 4.0 billion. The Veba Oel AG concept to use the same plant configuration for processing either coal or heavy oil or residue (but not in combination) allows more operating flexibility with changing market demand and economic requirements.

So far, no decision has been taken to proceed with these commercial plants. However, on the basis of past operating experience with I.G. Farben plants and the present state of development, it may be expected that the RAG/Veba process would reach commercial readiness by 1990.

2.3 RHEINBRAUN PROCESS

Rheinische Braunkohlenwerke AG (Rheinbraun) in collaboration with its subsidiary Union Rheinische Braunkohlen Kraftstoff AG is developing a direct, catalytic process for liquefaction of brown coal. Since 1978, they have been developing the process on a bench scale unit (250 kilograms per day of coal) at their laboratories at Wesseling, Rhein. The total expenditure of the bench-scale program (1978-86) is estimated to be U.S. \$3.0 million and is financed by the Federal Ministry for Research and Technology in Bonn (BMFT).

In the Rheinbraun process, a number modifications have been made over the classical Bergius/I.G.-Farben approach such as use of asphaltene-free distillate oil as recycle solvent, vacuum distillation and flashing methods for solid/liquid separation, and gasification of vacuum bottoms (residues) to produce make-up hydrogen for the process. These changes have resulted in increased coal throughput from 400 kg/h/m^3 to 600 kg/h/m^3, lowering of reactor pressure from 71 MPa to 30 MPa, reduction of reactor temperature from 478°C to 460°C, and lower hydrogen consumption (6.0 wt% instead of 6.5 wt% MAF coal).

While these improvements are similar to the RAG/Veba process described earlier, the Rheinbraun process incorporates some processing steps specifically for brown coals, such as:

- A proprietary method for drying of raw brown coal containing up to 60 wt% moisture.

- Thermal pretreatment of coal/oil/catalyst slurry at 350°C to remove residual moisture and to decompose the humic acids for prevention of calcium-carbonate induced deposits inside the preheater tubes and hydrogenation reactor.

2.3.1 History and Development of the Process

Rheinbraun and its subsidiaries were involved in the commercial-scale conversion of brown coal in the 1940's using the classical Bergius/I.G.-Farben approach. A coal liquefaction plant at Wesseling, south of Cologne, having an annual throughput of 3.75 million tonnes of raw brown coal was operated (1941-44) to produce about 2.0 million barrels per year of motor gasoline.

Following the war, efforts were mainly directed towards gasification of brown coal to produce synthesis gas for methanol manufacture, conversion of brown coals to briquettes for burning purposes, production of metallurgical coke, and low-CV fuel gas, etc.

In 1978, work was initiated on direct liquefaction of brown coals and similar low-rank coals to transportation fuels and/or feedstocks for chemical production. Several modifications, as described earlier, were incorporated in the new Rheinbraun process (also known as HVB* process). To verify these changes and to generate the experimental data, a bench-scale unit having a capacity of processing up to 250 kg/d of coal was set up at Rheinbraun facilities in Wesseling.

On the basis of these bench-scale experimental results and past operating experience (1941-44) of a commercial plant, Rheinbraun is now carrying out process design and engineering studies for a pilot plant which will process up to 360 t/d of raw brown coal. The expected target for start-up operation is projected for 1985. The ultimate objective is to develop the brown coal liquefaction technology to a commercialization stage by 1995.

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* HVB is an acronym in German for Hydrierende Verflüssigung von Braunkohle, meaning hydroliquefaction of brown coal.

2.3.1.1 Bench-Scale Program

A schematic diagram of the bench-scale unit at Wesseling is presented in Figure 2-8. Dried and pulverized coal is mixed with catalyst and recycle oil in a slurry preparation tank. Hydrogen is added to the slurry stream and the mixture is electrically heated to about 420°C before being fed into the bottom of an upflow, tubular reactor. The reactor operates at a temperature of 460°C and pressure of 30 MPa. The average residence time is about 30 minutes. The reactor effluents are cooled and separated into gases, liquids, and a residue slurry by conventional methods (flash separation, fractionation, vacuum distillation). Typical reactor operating conditions and product yields of Rhenish brown coal are summarized in Table 2-16. Analysis of the brown coal is given in Appendix B.

From Table 2-16, it may be seen that the operating conditions employed in the Rheinbraun coal liquefaction process are quite similar to the RAG/Veba process described earlier. The amount of hydrocarbon gases is 8 wt% MAF coal and the total liquid yield is about 46 wt% MAF coal. Hydrogen consumption is 6.0 wt% MAF coal.

As part of process development program, some laboratory tests on gasification of residual slurry have been made. A preliminary characterization of coal-derived liquids and an investigation of upgrading schemes have also been carried out. As stated by the process developer, the results obtained are encouraging.

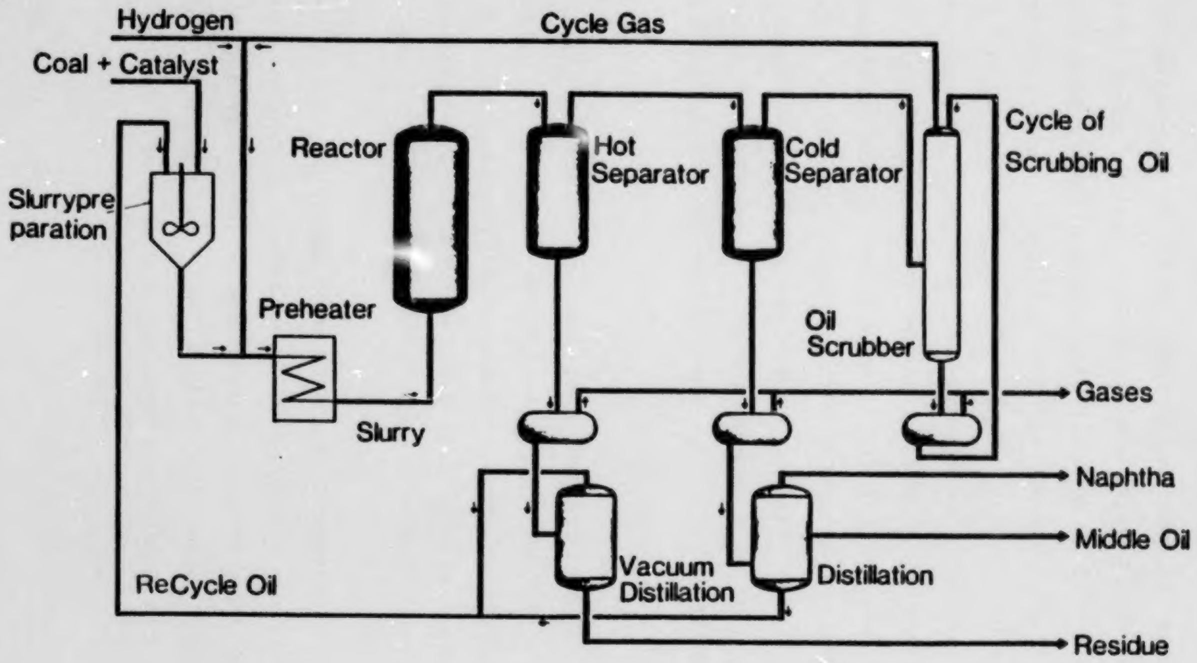

Figure 2-8 Schematic Diagram of the Rheinbraun Bench-Scale Unit at Wesseling (250 kg/d of coal)

Source: D. Bocker et al, Rheinische Braunkohlenwerke AG, "Rheinbraun Liquefaction of Brown Coal", paper presented at EPRI/KFA-Julich Conference on Synthetic Fuels, San Francisco, California, October 1980.

Table 2-16 Typical Operating Conditions and Product Yields obtained from Bench-Scale Experimental Unit at Rheinische Braunkohlenwerke AG (250 kg/d of coal)

Operating Conditions	
Coal Particle Size, mm	1.0
Coal Throughput, $kg/h/m^3$	600
Red Mud Consumption*, kg/t MAF coal	50
H_2 Partial Pressure, MPa	27
Preheater Outlet Temperature, °C	420
Reaction Temperature, °C	460
Total Reactor Pressure, MPa	30
Product Yields, wt% MAF coal	
Hydrocarbon Gases (C_1-C_4)	8.0
Naphtha (C_5-200°C)	13.8
Mid-Distillate (200-325°C)	32.2
Residue Slurry**	25.0
H_2O, NH_3, H_2S, CO_x	27.0
	106.0
Hydrogen Consumption	6.0
	100.0

* Red Mud contains 30 to 40 wt% Fe_2O_3 as catalytically active component.

** Residue slurry contains unreacted coal, catalyst and heavy oil (>325°C)

Source: D. Bocker et al., "Rheinbraun Liquefaction of Brown Coal", paper presented at EPRI/KFA-Julich Conference on Synthetic Fuels at San Francisco, California, October 1980.

2.3.2 **Technical Review**

2.3.2.1 Types of Coal Investigated

The new Rheinbraun process (HVB) has been developed specifically for the liquefaction of brown coal* from the Rhenish area, Germany. Low-rank coals from other countries such as Australia and New Zealand have also been tested.

The as-received Rhenish brown coal contains 60 wt% moisture and has an ash content of 3.2 wt% on MF basis. In comparison, the Estevan and Willow Bunch lignites contain 35 wt% moisture but the ash contents are high, i.e. 15.4 wt% for Estevan and 21 wt% for Willow Bunch lignites. In terms of their ultimate analyses on MAF basis, the Rhenish brown coal is comparable to Willow Bunch lignite as shown below.

	Rhenish Brown Coal	Estevan Lignite	Willow Bunch Lignite
Ultimate Analysis wt% MAF Coal			
Carbon	67.9	72.3	65.9
Hydrogen	5.1	5.3	5.1
Nitrogen	0.7	1.2	0.6
Sulfur	0.3	0.7	1.1
Oxygen	26.0	20.5	27.2

Proximate analysis of Rhenish brown coal is not available, hence, the comparison with Saskatchewan lignites is not possible. However, based on similarity of ultimate analysis of Rhenish brown coal and Estevan, it was assumed that the liquefaction behaviour of these coals are similar.

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* A typical analysis of Rhenish brown coal is given in Appendix B.

2.3.2.2 Overall Configuration of a Conceptual, Commercial Plant for the Rheinbraun Process

Figure 2-9 is a simplified block diagram of the Rheinbraun process. Raw coal from storage bins is fed into a crusher/dryer where the coal is pulverized to minus-0.1 mm and dried to about 10 to 12 wt% moisture. A major portion of the coal (67% of the total) is fed into a slurry preparation unit in the coal liquefaction section. The balance is used for hydrogen, fuel gas, steam and power generation.

The partially dried and pulverized coal is mixed with recycle oil (MF-coal/oil ratio of 0.75) and red mud (5 wt% MAF coal) to form a slurry containing 45 wt% solids. In the slurry preparation vessel, the temperature is maintained at 350°C to remove the residual moisture and to thermally decompose the humic acids present in low-rank coals. Thermal pretreatment is required in order to minimize the calcium-carbonate induced deposits in the furnace preheater tubes and hydrogenation reactor.

The high-pressure sour gas stream containing unreacted hydrogen is initially treated in a high-pressure scrubber to remove the acid gases and then cryogenically purified into a hydrogen-rich stream which is recycled to the reactor. The remaining treated gas (sweet gas) containing mainly lighter hydrocarbons (C_1-C_4) is sent for LPG recovery. Low-pressure sour gas recovered from flash separation, fractionation and naphtha stabilization is also sent for gas treatment, and the treated gas is mainly used as fuel gas for the process. Acid gases are sent to the Claus plant for sulfur recovery.

The condensed liquid hydrocarbon stream is preheated and then fractionated in the atmospheric distillation tower to produce primary distillate products: naphtha, mid-distillate or middle oil, and heavy oil. A portion of middle oil and all the heavy oil is recycled to the coal-oil slurry preparation unit.

The solids-containing slurry is fed into the vacuum distillation tower to recover a portion of distillate oil which is recycled to the coal-oil slurry preparation unit. The vacuum tower bottoms (residual slurry) containing about 50 wt% solids (unreacted coal, ash and catalyst) are pumped to the hydrogen production section.

Sour water containing crude-mixed phenols and dissolved ammonium salts is sent to the effluent control unit for recovery of phenol and ammonia. Acid gases are sent to the Claus plant for sulfur recovery.

Most of the steam is generated by recovering process heat and the balance is produced by burning coal fines. Electricity is generated on site by burning coal.

2.3.2.3 Product Yield and Efficiencies

The product yield and distribution of a conceptual, self-sufficient, commercial plant using the Rheinbraun process is presented in Table 2-17.

The net liquid product (C_5-325°C) yield is about 0.7 bbl/t of as-received Rhenish brown coal compared to 1.1 bbl/t of Estevan lignite. In other words, the yield on an as-received basis for Rhenish brown coal is 36% less because of its high moisture content (60 wt% H_2O). On MF basis, the yield for Estevan lignite

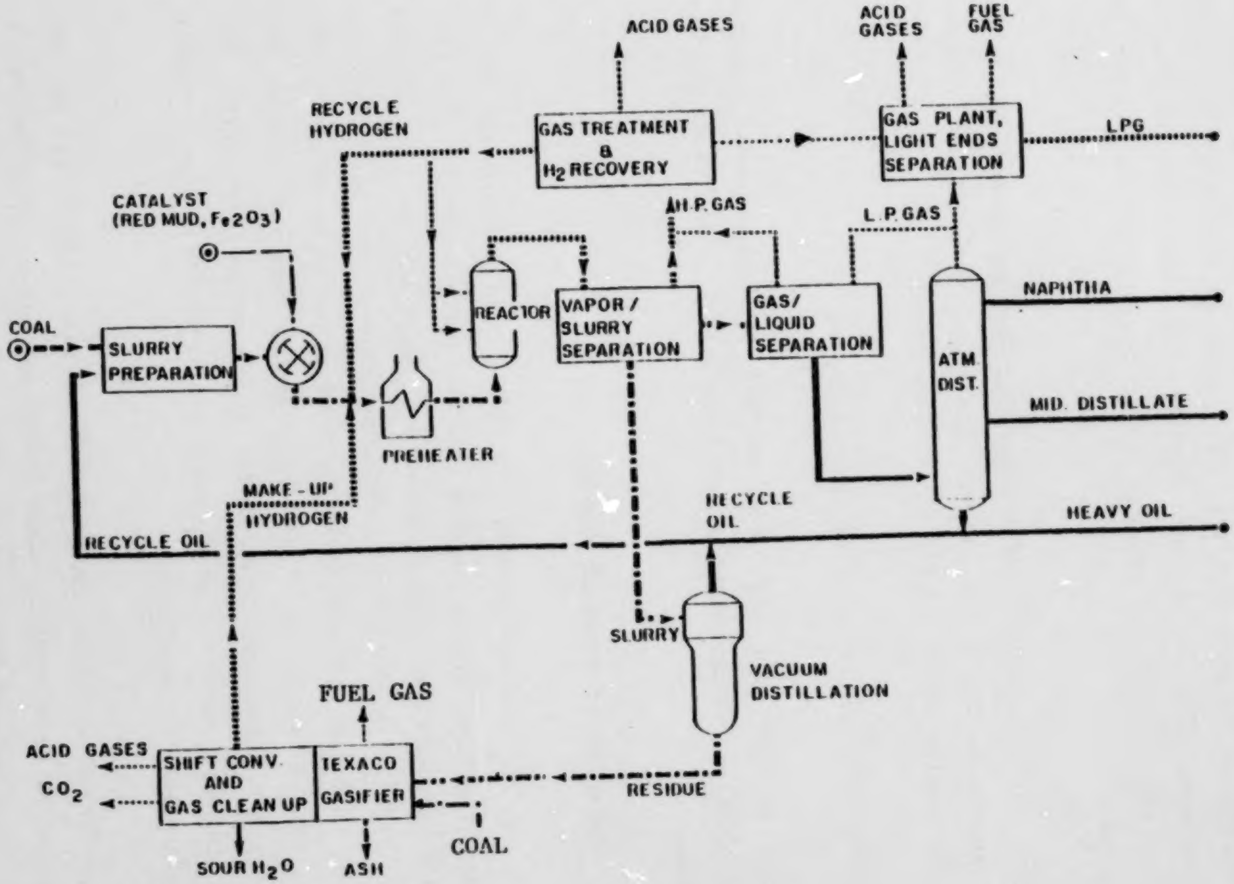


Figure 2-9 Simplified Diagram of the Rheinbraun Process

is lower than for Rhenish brown coal because of the higher ash content of Estevan lignite. The yields on MAF basis are almost the same. For Willow Bunch lignite, the liquid product yield on MAF basis would be somewhat lower because of the higher ash content.

The "overall product" efficiency for Rhenish brown coal is 51% compared to 52.2% for Estevan lignite. The "liquid product" efficiencies are 44.1% and 43.2% respectively.

2.3.2.4 Suitability/Availability of Process Inputs

High moisture and high ash contents of Saskatchewan lignites would affect their behaviour and processibility in the Rheinbraun process.

- **Reduction of Moisture Content**

The initial moisture content (35 wt%) of the Estevan and Willow Bunch lignites may be reduced to the desired level (<2 wt%) using a two-stage drying technique suggested by the process developer. Initially, the moisture content may be reduced to about 10 wt% by air drying in the coal preparation section. The residual moisture is then subsequently removed during the "thermal pretreatment" step in the slurry preparation unit. The second stage employs a proprietary slurry-drying method, details of which are not available.

Reduction of moisture from 35 wt% to less than 2 wt% is not expected to be a problem, since Rheinbraun has acquired extensive experience and has demonstrated that brown coals containing up to

Table 2-17 Product Yield and Efficiencies of the Rheinbraun Conceptual Plant

(Basis: 25,000 t/d as-received coal)

	Rhenish Brown Coal	Estevan Lignite
Coal (as received)		
Moisture, wt%	60.0	35.0
Ash, wt% of MF coal	3.2	15.4
Input		
MF Coal to Reactor, t/d	6,535	10,860
MF Coal to Gasifiers* t/d	660	1,555
MF Coal to Utilities**, t/d	2,805	3,835
Electricity, MW	-	-
Output t/d (bbl/d)		
SNG	-	-
LPG	253	592
Naphtha (C_5 - 200°C)	756 (5,950)	1,323 (10,410)
Mid-Distillate (200 - 325°C)	1,727 (12,278)	2,397 (17,040)
Heavy Oil (325 - 525°)	-	-
Sulfur	27	56
Phenol	-	-
Ammonia	40	76
Liquid Product Yield (C_5-325°C)		
bbl/t as-received coal to plant	0.7	1.1
bbl/t MF coal to plant	1.8	1.7
bbl/t MAF coal to plant	1.9	2.0
Naphtha in Total Liquid, wt%	30.0	35.6
Thermal Efficiency, %		
Overall	51.0	52.2
Liquid Products	44.1	43.2

* Coal for fuel gas and hydrogen production

** Coal for steam and power generation

Source: D. Bocker et al., "Rheinbraun Liquefaction of Brown Coal", paper presented at EPRI/KFA-Julich Conference on Synthetic Fuels at San Francisco, California, October 1980.

60 wt% moisture may be dried to the desired level without significantly affecting the properties of the feed coal and the product yield.

- **Ash Content of Saskatchewan Lignites**

The Estevan and Willow Bunch lignites contain 15.4 wt% and 21 wt% ash respectively compared to 3.2 wt% in Rhenish brown coal. The high ash content of Saskatchewan lignites together with the addition of red mud catalyst ($\simeq$5 wt% of MAF coal feed) to the reactor system may contribute to operational difficulties associated with the pumping and handling of the highly viscous and abrasive-solids-containing residual slurry.

In the conceptual plant configuration for Estevan lignite, about 4290 bbl/d of recoverable oil was allowed to be lost in the vacuum-bottoms streams (residual slurry) so as not to exceed the limit of 50 wt% solids. The net liquid yield decreased by 0.4 bbl/t of MF coal, and the "liquid product" efficiency dropped from 49% to 43% in comparison to processing of Rhein brown coal. For Willow Bunch lignite, having 21 wt% ash (MF basis), the decrease in net liquid yield would be even more than for the Estevan lignite.

- **Availability of Process Inputs**

In the Rheinbraun process, the main process inputs are coal, water and catalyst. Requirements of water and red mud catalyst are 525 L/s and 168,330 tonnes per year respectively. Electricity is generated onsite.

2.3.2.5 Product Quality, Utilization and Upgrading Schemes

Preliminary evaluation of product quality, utilization and upgrading schemes for coal-derived liquids obtained from bench-scale processing of Rhein brown coal have been carried out. Analysis and characterization of the liquid products are not available.

According to Rheinbraun information, the naphtha requires catalytic hydrotreating and reforming to produce reformate suitable as blendstock for motor gasoline. During hydrotreatment, a pressure of 10 MPa is required for removal of heteroatom impurities. The unleaded reformed gasoline has a research octane number (RON) of 97. The middle oil requires severe hydrotreatment for removal of heteroatom impurities. A pressure of 20 to 30 MPa is required. The hydrotreated product has a Cetane number of 50 and meets the diesel fuel requirements.

2.3.2.6 Areas Requiring Further Development

The overall processing scheme and operating parameters of the Rheinbraun process is quite similar to the RAG/Veba process described earlier. Therefore, in general terms, the problem areas pertaining to the RAG/Veba process would also apply to the Rheinbraun process. In addition, testing of the two-stage proprietory coal drying method and the thermal pretreatment step would be required on a large scale. These unit operations should be demonstrated in the proposed pilot plant.

2.3.2.7 Status and Commercialization Prospects

The new Rheinbraun process (HVB) has been under development since 1978. So far, the process has been tested in a bench-scale unit having a capacity of 250 kg/d of coal. Most of the coal liquefaction experiments have been carried out on Rhenish brown coal, and sufficient new experimental data supplemented with past operating experience (1941-44) of the large-scale plant has been accumulated to permit a scale-up of the new Rheinbraun technology to the pilot-plant scale (360 t/d of raw brown coal). As part of the bench-scale program, preliminary tests on gasification of vacuum bottoms (residual slurry) for production of make-up hydrogen have been made. The characterization of coal-derived products and an investigation of upgrading schemes to produce transportation fuels are in progress.

Since 1980, Rheinbraun is engaged in the detailed engineering and process design of a 360-t/d pilot plant. The mechanical completion of the pilot plant and the start-up operation are scheduled for 1985, but the target appears to be optimistic. The ultimate objective is to build a single-train commercial plant (annual raw brown coal throughput of 3.0 million tonnes) by 1992 and a full-scale plant (annual raw brown coal throughput of 20 million tonnes) by 1995.

2.4 EXXON DONOR SOLVENT PROCESS

The Exxon Donor Solvent (EDS) coal liquefaction process converts a wide varity of coals into distillate hydrocarbons. The main feature of the process is the role of the recycle solvent which donates hydrogen to the coal during liquefaction.

The other features of the EDS process are:

- Crushed coal is liquefied in a non-catalytic tubular reactor in the presence of molecular hydrogen and a hydrogen donor solvent.
- The donor solvent is a 205 to 545°C boiling-range material which is hydrogenated in a separate fixed-bed catalytic hydrogenation step.
- The liquefaction reactor operates at 405 to 470°C temperature and 10.3 to 13.8 MPa pressure and the reactor effluents are separated by distillation into gas, naphtha, distillates, and a vacuum-bottoms slurry.

2.4.1 History and Development of the Process

The development and demonstration of the process, which began in 1968, has been carried out in microautoclaves, bench-scale units ranging in size from 7 to 45 kilograms per day and in two process development units (PDU) handling 0.5 t/d and 1 t/d of coal respectively.

The majority of the concepts and innovations were originated in the 0.5 t/d Coal Liquefaction Pilot Plant (CLPP-1). Based on the results of bench-scale research and PDU operations, detailed engineering, procurement, and construction of a 227-t/d pilot plant began in mid-1977 and was completed in March 1980 at a total cost of $118 million. The plant was commissioned in mid-1980 and is presently in operation.

The overall project objective is to achieve commercial readiness for the EDS process by 1983. This should be accomplished through the following more specific objectives:

- Develop the EDS process and the design bases necessary to liquefy a range of different coal ranks in an environmentally acceptable manner.
- Develop process correlations to predict product yields and quality for different operating conditions and feed coals.
- Estimate economics of a commercial plant.
- Conduct a product testing program to ensure marketable products.

A cooperative agreement to accomplish these objectives was signed between the industry and the U.S. Department of Energy (DOE) in 1977. The first 10 years of the pre-development work on the EDS process from 1966 to 1976 was completely financed by Exxon, U.S.A. Since the inception of the joint venture, DOE is providing 48% of the $437 million in the EDS program. The balance is funded by the following seven industrial participants:

- Exxon Coal Company U.S.A. ($81 million)
- Electric Power Research Institute ($42 million)
- Japan Coal Liquefaction Development Co., Ltd. ($27 million)
- Phillips Coal Company ($7 million)

2.4 EXXON DONOR SOLVENT PROCESS (cont'd)

- ARCO Coal Company ($7 million)
- Ruhrkohle A.G., Germany ($7 million)
- Agip S.p.A., Italy ($7 million)

2.4.1.1 Bench-Scale and PDU Program

In the development of the EDS Process, extensive use has been made of small, integrated coal liquefaction bench-scale and process development units of 45 kg/d and 1 t/d feed coal capacities. The key features of these units are shown schematically in Figure 2-10. In the slurry preparation area, coal is slurried with a recycle donor solvent. In the smaller unit, a batch preparation technique with manual addition of the solvent, crushed coal and bottoms (if recycled) at six-hour intervals is used. In the larger 1-t/d unit, a continuous slurry preparation method with metering of coal, solvent, and bottoms is used. In both units, feed slurry is pumped to the liquefaction reactor from a slurry holding tank using high-pressure positive displacement pumps.

The slurry is mixed with hydrogen before preheating and fed to tubular upflow reactors in both cases. Several reactors in series may be connected by tubular transfer lines to achieve the desired residence time.

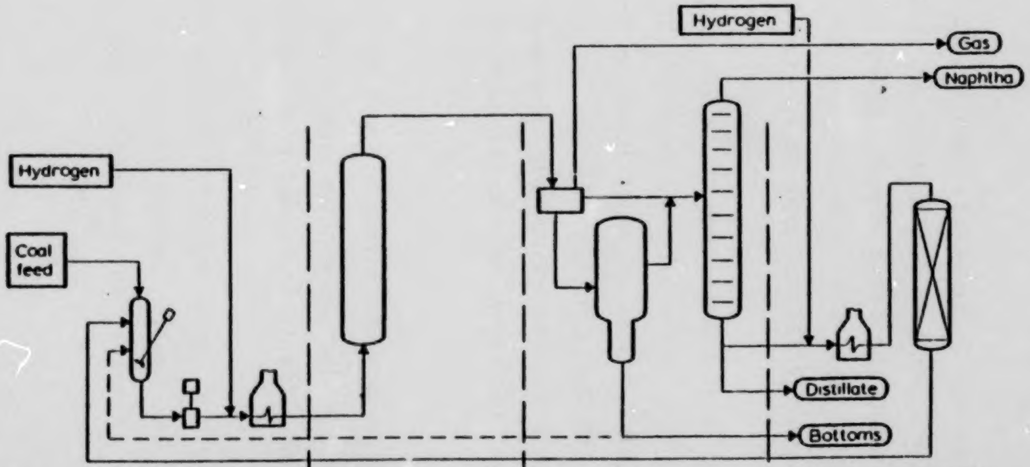
The effluents leaving the liquefaction reactor are separated by conventional methods. Gases are separated by high and low pressure flashes. The unconverted coal and mineral matter are separated from the heavier coal-derived liquids by gas stripping in the smaller unit and by vacuum distillation in the 1-t/d unit. Products are split into naphtha and distillate using fractionation towers in both cases. A portion of the distillate fraction is used for producing the recycle solvent.

2.4 EXXON DONOR SOLVENT PROCESS (cont'd)

This unhydrogenated distillate is introduced along with molecular hydrogen into a conventional fixed-bed catalytic reactor. The operating conditions are adjusted to produce a recycle solvent of desired quality and specifications, which are defined by a single parameter termed as Solvent Quality Index (SQI). The SQI characterizes the hydrogen donor ability of the solvent, details of which are proprietory.

Bituminous, subbituminous and lignitic coals have been extensively tested in these units to identify and determine the effect of process variables on the liquefaction yields in a 'once-through liquefaction' mode of operation. These variables include reactor residence time, temperature, pressure, solvent-to-coal ratio and the Solvent Quality Index. In the 'bottoms-recycle liquefaction' mode of operation, the effect of partial recycle of liquefaction residue (vacuum bottoms) on coal conversion and product yields has also been investigated. The range of process variables studied to determine the liquefaction response from lignitic coals (North Dakota from Indian Head and Texas lignite from the Big Brown mine) are given below. Analyses of these coals is given in Appendix B.

Temperature = 427 to 454°C
Nominal Slurry Residence Time = 40,60 and 100 min
Hydrogen Partial Pressure in the Reactor = 10.3 to 13.8 MPa
Hydrogen Treat Rate = 4.5 to 9.5 wt% MAF coal
Solvent-to-Coal Ratio = 1.6 to 2.8
Recycle-Bottoms-to-Coal Ratio = 0.5 to 1.0



Unit	Key Features	Slurry preparation	Liquefaction	Product separation	Solvent hydrogenation
Bench Scale	45 kg/d	Batch slurry prep'n	Tubular upflow reactors	Bottoms sep'n by gas stripping	Fixed-bed cat reactor
PDU	1 t/d	Continuous slurry prep'n	Tubular upflow reactors	Bottoms sep'n by vacuum distillation	Fixed-bed cat reactor

Figure 2-10 EDS Bench-Scale and Process Development Units

Source: J.W. Taunton et al., FUEL, Vol. 60, September 1981

The liquefaction response for North Dakota lignite, expressed as wt% MAF coal, for C_4-525°C product is found to be about 29 wt% compared to 33 wt% obtained for Texas lignite under the similar operating conditions (10.3 MPa, 449°C, H_2 treat-rate 4 wt% MAF coal, SQI of 4, and nominal residence time of 40 minutes).

When the residence time is increased from 40 to 100 minutes, the C_4-525°C product yields from Texas lignite decreases by 4 wt% on MAF coal basis whereas, under similar conditions, an increase of about 2 wt% is found with North Dakota lignite. At 100 minutes residence time, the Texas lignite presumably undergoes cracking and the C_1-C_3 gas make increases at the expense of C_4-525°C product yields.

The effect of a hydrogen treat-rate beyond 4.5 wt% of MAF coal is insignificant on liquid yields. Also, the yields are the same when solvent-to-coal feed ratio is increased from 1.6 to 2.0.

The hydrogen consumption for Texas and North Dakota lignites is about 4.0 and 5.0 wt% MAF coal, respectively. The lower H_2 consumption for Texas lignite is primarily due to lower conversion.

The amount of sour gases produced (CO_x, H_2S) and the water formation for both lignites are found consistent with their elemental analysis.

Some typical liquefaction yields are presented in Table 2-18. Amongst the lignitic coals, only Texas lignite has been tested in the "bottoms-recycle" mode. The results indicate an increase of about 7 wt% in liquid yields when 50% of the bottoms are recycled back to the liquefaction reactor. However, this increase is achieved at 427°C, 13.8 MPa and a 100 minute nominal residence time.

2.4.1.2 Pilot Plant Program

The Exxon Coal Liquefaction Pilot Plant (ECLP) is designed to process 227 tonnes of coal per day. The plant is located near Exxon's refinery at Baytown, Texas, and was commissioned in June 1980.

A schematic flow plan of the EDS pilot plant (ECLP) is shown in Figure 2-11. Raw coal is dried and crushed in either a gas-swept rotary mill or crushed in an impact mill and subsequently dried in a slurry drier. When operating on prepared, dried and crushed coal, the slurry drier serves as a mixer of the recycle hydrogen-rich donor solvent and the coal. When operating on crushed as-received coal, it serves as a drier as well as a mixer.

The coal/oil slurry is pumped to over 13.8 MPa pressure, gaseous hydrogen is added, and the three-phase mixture is heated to about 449°C in one of two slurry preheat furnaces. The second furnace serves as a spare if it is necessary to decoke the first furnace. The heated mixture is then passed upflow through the liquefaction reactor system. There are four reactors available

Table 2-18 Liquefaction Yields from Lignites in the EDS Process

Coal	North Dakota "Indian Head"	Texas "Big Brown"		
EDS Run No. (Process Unit)	415 (RCLU)	532-536 (RCLU)	344 (CLPP)	338-340 (CLPP)
Liquefaction Conditions				
Temperature, ^{o}C	449	449	449	427
Hydrogen Partial Pressure, MPa	10.3	10.3	13.8	13.8
Nominal Residence Time, min.	40	40	60	100
Solvent-to-Coal Ratio	1.6	1.6	1.6	1.6
Solvent Quality Index	4.0	4.3	4.2	N.A.
Hydrogen Treat Rate, wt% MAF coal	4.4	4.8	4.8	7.2
Bottoms-to-Coal Ratio	0	0	0	0.5
Overall Yields, wt% MAF coal				
Water Carbon Oxid	20.57	13.04	13.64	13.96
C_1-C_3 Gases	8.09	9.09	8.78	9.09
Naphtha (C_4-200^{o}C)	20.15	20.57	26.36	28.02
Oil (200-525^{o}C)	9.30	12.80	12.76	18.58
Total C_4-525^{o}C Liquid	29.45	33.37	39.12	46.60
> 525 °C Bottoms	38.46	41.25	39.69	48.78
Hydrogen Consumption	4.9	3.83	4.32	5.14

Notes:
* - In the product yield data, EDS has included butane (C_4) in naphtha fraction
RCLU - Recycle Coal Liquefaction Unit (45 kg/d of coal)
CLPP - Coal Liquefaction Pilot Plant (1 t/d of coal)

Sources: DOE Reports FE-2353-20, Feb.78; FE-2893-21, Dec.81; FE-2893-70 (Vol.I), July 81

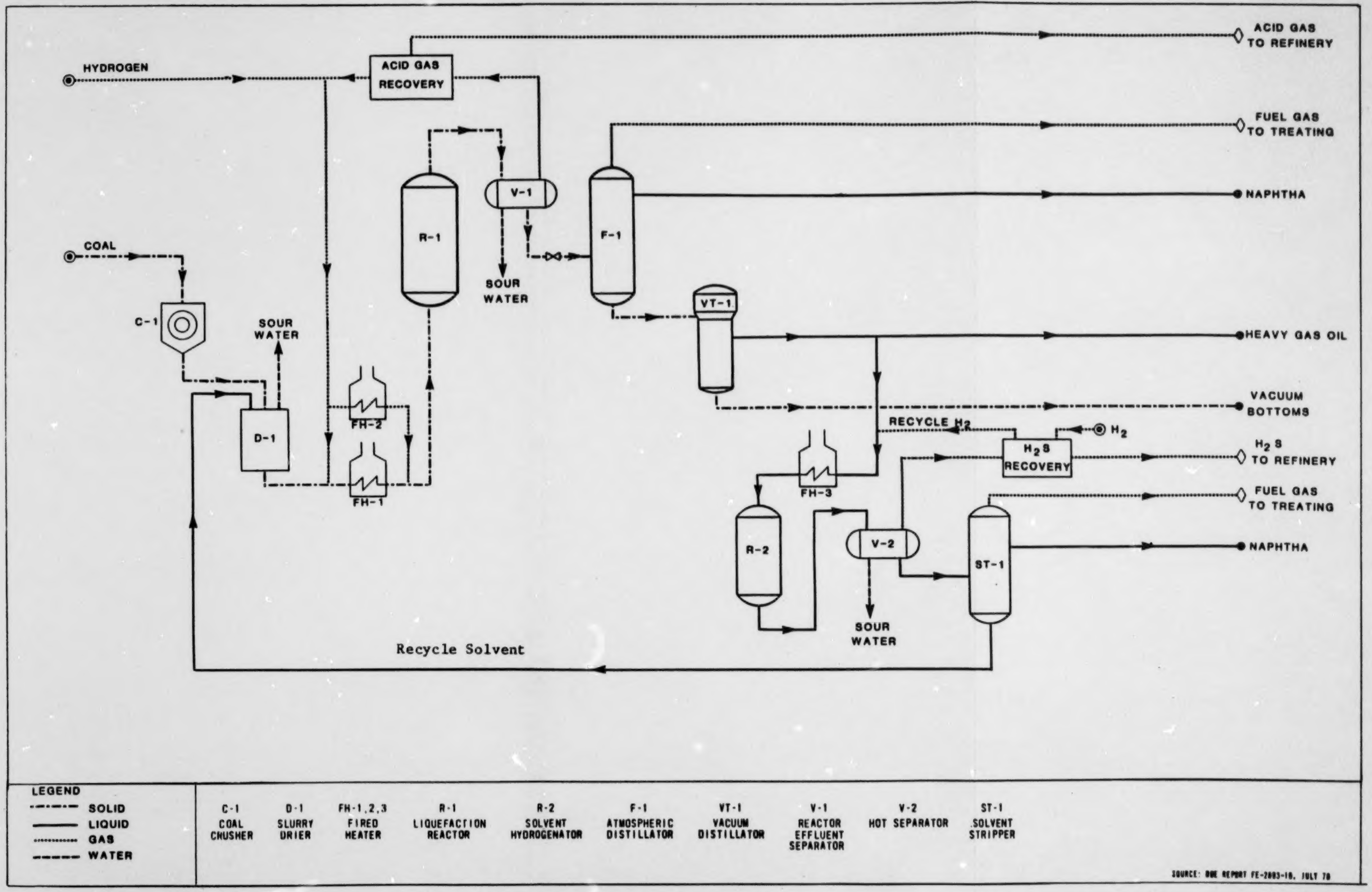


Figure 2-11 Schematic Flow Plan of the 227-tonnes-per-day Exxon Coal Liquefaction Pilot Plant (ECLP) at Baytown, Texas

Source: DOE Report FE-2893-16, July 1978

at ECLP. During the Illinois No.6 coal operation, only two have been used to provide the desired residence time of 40 to 60 minutes.

The effluent from the reactors passes through a series of separator drums into an atmospheric fractionator where the naphtha is taken out as an overhead product, sidestreams of gas oil are withdrawn and the atmospheric bottoms are pumped through a preheat furnace to a vacuum distillation tower and heavy gas-oil is withdrawn as a side stream from the vacuum tower. The vacuum bottoms from the tower are solidified.

A portion of the vacuum gas-oil is sent to a solvent hydrogenation unit where hydrogen is added in a fixed-bed catalytic reactor to increase the hydrogen donor capability of the solvent. Four solvent hydrogenation reactors are used at ECLP. The hydrogenated product is fractionated to separate the solvent for recycle to the slurry drier. Naphtha and excess gas oil along with heavy-vacuum gas-oil are sent to storage tanks.

The bottoms from the vacuum tower are solidified, collected in bins and stored for testing in the "bottoms processing" program. The overall ECLP configuration represents an integration of several interrelated processing units. The successful operation of ECLP will therefore demonstrate that these units can be operated together in an integrated manner over a sustained period of time.

The first integrated ECLP run was made on June 24, 1980 with Illinois No.6 coal. The longest run to date has been of 60 days duration with a coal-in factor of 47% versus a target of 50%. The 11 months of operation on an Illinois bituminous coal was completed on June 2, 1981. The tests, including a five-month shakedown run and a six-month formal test program, totaled more than 3,900 hours. Since mid-July '81, the plant is operating on Wyoming coal, including the testing of vacuum bottoms recycling. After the testing program on Wyodak coal will be completed, Texas lignite is planned for testing.

2.4.2 **Technical Review**

Presented below is a technical review of the EDS process based on the information available in the open literature and material related to the processing of lignitic coal directly received from the process developer.

2.4.2.1 Types of Coals Investigated

Different types and ranks of coal have been processed on bench-scale and PDU units. Table 2-19 presents the breakdown of the accumulated number of hours spent to determine yields on these units. Amongst the coals tested, over 10,000 hours in more than 40 runs have been logged with bituminous Illinois No.6 coal. The longest run was 610 hours in the bench-scale unit (RCLU) and 1440 hours in the 227-t/d pilot plant. ER&E has made an engineering design for a conceptual commercial plant for this coal. Sub-bituminous Wyodak coal has been investigated for over 6000 hours in more than 50 runs. The longest run in RCLU was 171 hours. The engineering design of a conceptual, commercial plant for Wyodak subbituminous coal is underway.

Amongst the lignitic coals, Texas Big Brown lignite has been processed for about 10,000 hours in RCLU with a longest run of 139 hours. This coal is scheduled for 6-month processing in the 227-t/d pilot plant.

Table 2-19 Accumulated Hours of EDS Bench-Scale and PDU Runs

Type of Coal	Origin	Mine	No. of Hours	Total Hours
Bituminous	Pittsburgh	Ireland	1,332	
		Arkwright	320	
	Illinois	Monterey No.1	13,900*	
		Burning Star No. 2	833	16,385
Subbituminous	Wyoming	Wyodak	6,040	
	Australian	Wandoan	845	
				6,885
Lignite	Texas	Big Brown	998	
	North Dakota	Indian Head	470	
				1,468

* Include 3,900 hours on 227 tonne-per-day pilot plant

2.4.2.2 Overall Process Configuration of the Conceptual, Commercial Plant of the EDS Process

A simplified conceptual EDS process configuration is shown in Figure 2-12. A conceptual, self-sufficient, integrated, commercial EDS plant when FLEXICOKING is used (so-called Design-Update Case) is presented. A brief description of the main sections are given below.

- **Coal Storage and Preparation**

The coal conversion plant receives cleaned coal from the mines via conveyor belts. The coal is stored in a stacker/reclaimer area. The cleaned coal feed to the liquefaction plant is reduced from 90% minus 25-mm to minus 2.5-mm in impact-mill crushers and then fed to the slurry preparation block.

- **Coal Liquefaction, Product Recovery and Fractionation**

Crushed coal is fed to slurry driers through enclosed belt conveyors and is slurried with recycle donor solvent. The coal in the slurry is dried to less than 4 wt% moisture, pumped to reaction pressure, mixed with hydrogen treat gas and then preheated in the fuel-fired furnace. The mixture is then fed to the liquefaction reactor operating at 449°C and 13.8 MPa.

Products from the liquefaction reactor are separated into a gas stream and a liquid/solids stream in the vapor/slurry separation block. The gas stream is cooled to separate the vaporized naphtha and solvent and then scrubbed with diethanolamine (DEA) for removal of NH_3, H_2H and CO_2.

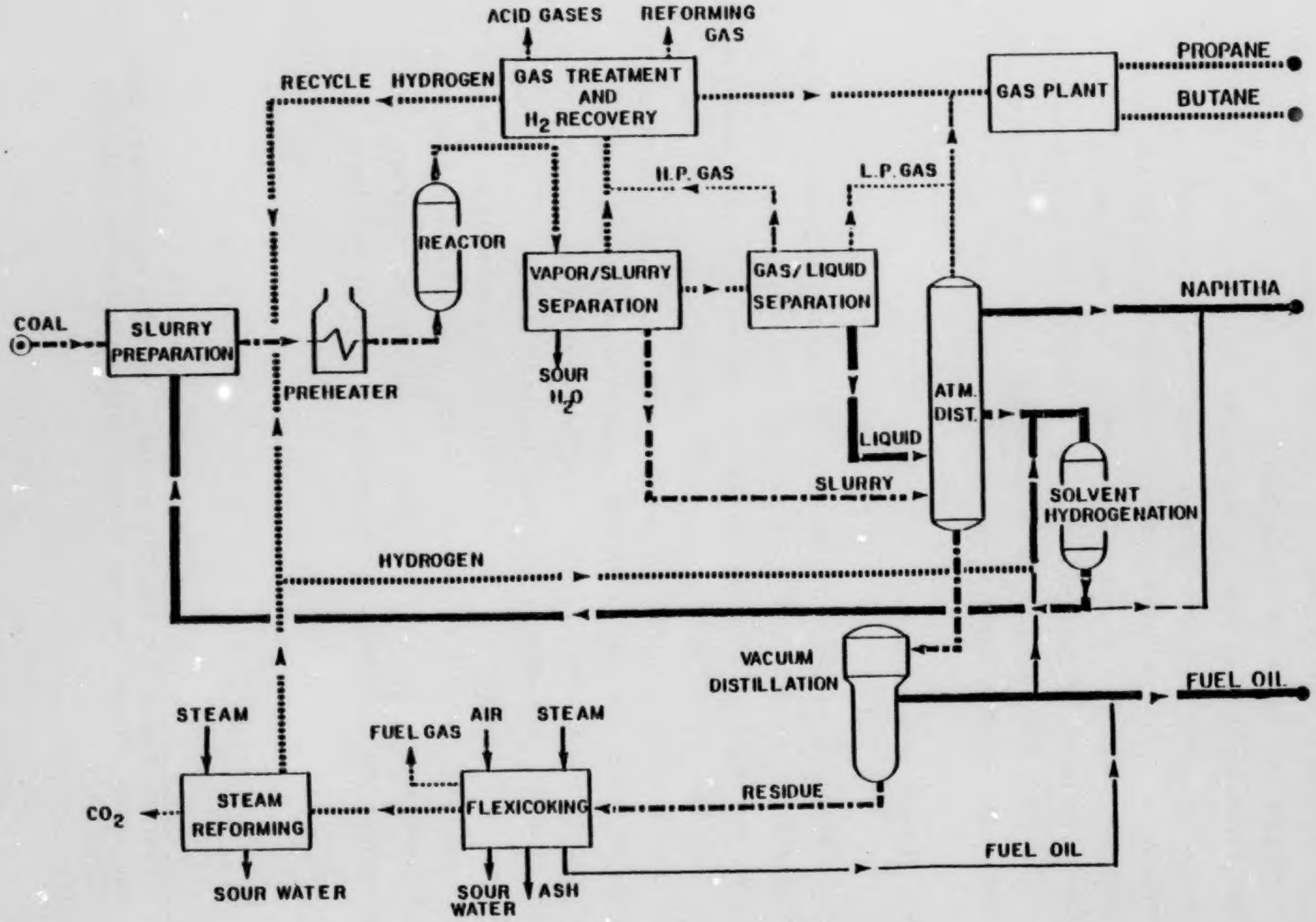


Figure 2-12 Overall Conceptual EDS Process Configuration

The liquid/solids stream is fed to atmospheric and vacuum distillation units to fractionate the liquefaction products. Recovered naphtha is sent to light-ends processing for stabilization. A spent solvent stream consisting of atmospheric and vacuum fractionator sidestreams is recycled to the solvent hydrogenation unit. Vacuum gas-oil product is removed from the vacuum tower and sent to fuel-oil product storage. The vacuum bottoms ($>$525°C) is fed to the FLEXICOKING unit for recovery of additional liquids and fuel-gas manufacture.

- **Solvent Hydrogenation**

The processing sequence for solvent hydrogenation closely resembles that of petroleum hydrofining systems. The spent solvent is hydrotreated in a fixed-bed catalytic reactor. The unit operates at 12.3 MPa and at an average reactor temperature of RV* + 52°C. Effluents from the reactor are cooled and separated into a hydrogen-rich gas and a hydrotreated liquid stream. The gas after NH_3 and H_2S removal is recycled back to the reactor along with make-up hydrogen. The hydrotreated liquid is fed to a solvent fractionator. In addition to hydrogen-donor recycle solvent, this unit also produces a hydrogenated low-sulfur fuel oil product, naphtha which is sent for light-ends recovery, and a gas stream which is sent to the hydrogen recovery section.

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* Reference Value (In the EDS process, certain proprietary data are coded through the use of confidential reference values).

- **Flexicoking**

In the FLEXICOKING unit, liquid and gas products are recovered and coke is formed from the vacuum bottoms slurry. The FLEXICOKING unit consists of three vessels: fluid-bed coker reactor, heater and gasifier. The vacuum bottoms are fed to the reactor where the heat is supplied by the coke stream from the heater. The heater is in turn kept hot by circulating gases and solids from the gasifier where the coke formed in the reactor is gasified with steam and air to form low-CV gas (CV of 4.7 MJ/m^3).

Liquid and gas products are cooled and scrubbed and a condensed solids-laden stream is withdrawn. The solids content in this stream is reduced by hydroclones before it is blended with the product fuel oil. The olefinic coker gas stream is sent to the hydrogen recovery unit after acid gas removal.

- **Hydrogen Recovery/Generation and Ammonia Synthesis**

Hydrogen recovery and generation facilities include steam reforming of C_1-C_3 for hydrogen generation, and cryogenic purificaton of purge and offgases for recovery of hydrogen, C_1-C_3, and C_3+. The processing sequence and all equipment included in the hydrogen facilities (with exception of the stream reformer furnaces) represent established commercial technology. The reformer furnaces are somewhat unconventional in that the primary fuel is a low-Btu gas. The synthesis gas from the steam reformer furnace is upgraded in a conventional manner using two stages of high temperature shift reactors followed by acid gas removal and methanation. After methanation, the hydrogen stream is compressed. The ammonia synthesis is carried out downstream from hydrogen compression to prevent nitrogen gas build-up in the recycle gas system.

- **Light Ends Recovery**

The naphtha stream from liquefaction, solvent hydrogenation and FLEXICOKING are blended with condensate from the cryogenic hydrogen recovery section and fed to a conventional light-ends system. A deethanizer strips out the C_2^- fraction, which is sent to steam reforming as feed, and the bottoms are fed to a debutanizer. The overhead from the debutanizer is treated for sulfur removal and fed to a C_3/C_4 splitter to separate propane and butane as LPG products. The debutanizer bottoms is a stabilized naphtha (C_5-200°C) product.

- **Emission and Effluent Control Systems**

The EDS plant includes sour-water treatment facilities where ammonia is recovered in anhydrous form as a byproduct. It also includes a sulfur and a tail-gas treatment plant. The feed to the Claus sulfur plant consists of stripped H_2S and CO_2 gases from the DEA regeneration unit. About 99.9% of the sulfur is recovered as byproduct and thus the SO_2 emission to the atmosphere is reduced to about 100 ppm level.

2.4.2.3 Alternative EDS Process Configurations

Exxon Research and Engineering Company (ER&E), the EDS process developer, has also examined several other process alternatives.

In particular:

- "Two-Stage Liquefaction Scheme" with the objective to increase coal conversion. Under this alternative, bottoms from the first liquefaction reactor are fed directly to a second-stage reactor which operates at a 1.4 MPa lower pressure compared to the first reactor. The conversion increases by 5 wt%; however, an economic evaluation for a "commercial plant setting" indicated that costs would exceed by $80 million per year the credits gained by increased conversion.

- "Market Flexibility Sensitivity Case" with the objective to produce a salable medium-CV fuel gas and to reduce plant overall fuel and steam consumption in the plant. Under this configuration, only half of the vacuum-bottoms residue is fed into the FLEXICOKER and the remaining residue is separately gasified in a partial oxidation unit. Compared to the "Design Update Case", naphtha and fuel oil yields are nearly the same. In addition, about 65 kg of fuel gas per tonne of dry coal is also produced. However, the electricity requirement from outside sources increases by about 38%.

- "Vacuum-Bottoms Recycle Case" with the objective to achieve additional conversion of liquefaction bottoms residue. Under this configuration, a part of the vacuum bottom residue is recycled to the liquefaction reactor which is now operated at a longer residence time (60 min). The nominal hydrogen partial pressure in the reactor is also higher (13.8 MPa as compared to 10.3 MPa for Illinois coal). The equipment cost for the coal liquefaction and solvent hydrogenation units increases by 62% because of the recycling of vacuum bottoms (a solvent/coal/bottoms ratio of 2.5/1.0/1.0 is maintained for Illinois coal).

- "Alternatives for Hydrogen and Fuel Production" include the following configurations:

Hydrogen	Fuel	Steam
Partial oxidation of coal in a moving bed	Partial oxidation of coal	Vacuum-bottoms fired boiler
Steam reforming C_1/C_2 gases	Direct combustion of vacuum bottoms	Coal-fired boiler
Partial oxidaton of vacuum bottoms either by Texaco or Shell-Koppers gasifier	Gasification of vacuum bottoms by Texaco or Shell-Koppers gasifier, or combustion of C_1/C_2 gas	C_1/C_2 gas-fired boiler

The above alternatives were evaluated using a computerized linear programming model called the "Process Alternative Model" (PAM). The model is based on "Base Design" process configuration data and consider processing costs and product yields as well as restrictions due to material balance, feed availability, product demands, etc. The results of the PAM study which was recently conducted for Texas lignite are presented in Figure 2-13. These results indicate that the relative production cost of the coal-derived liquids is sensitive to the choice of bottoms processing configuration. The base case for the comparison is steam-reforming/FLEXICOKING configuration. When hydrogen is provided by partial oxidation of vacuum bottoms and process fuel is provided by the partial oxidation of coal in a hybrid boiler, the product cost of liquids is about 25% lower than the base case. The "bottoms recycle" mode is also being evaluated at a 227 t/d pilot plant level.

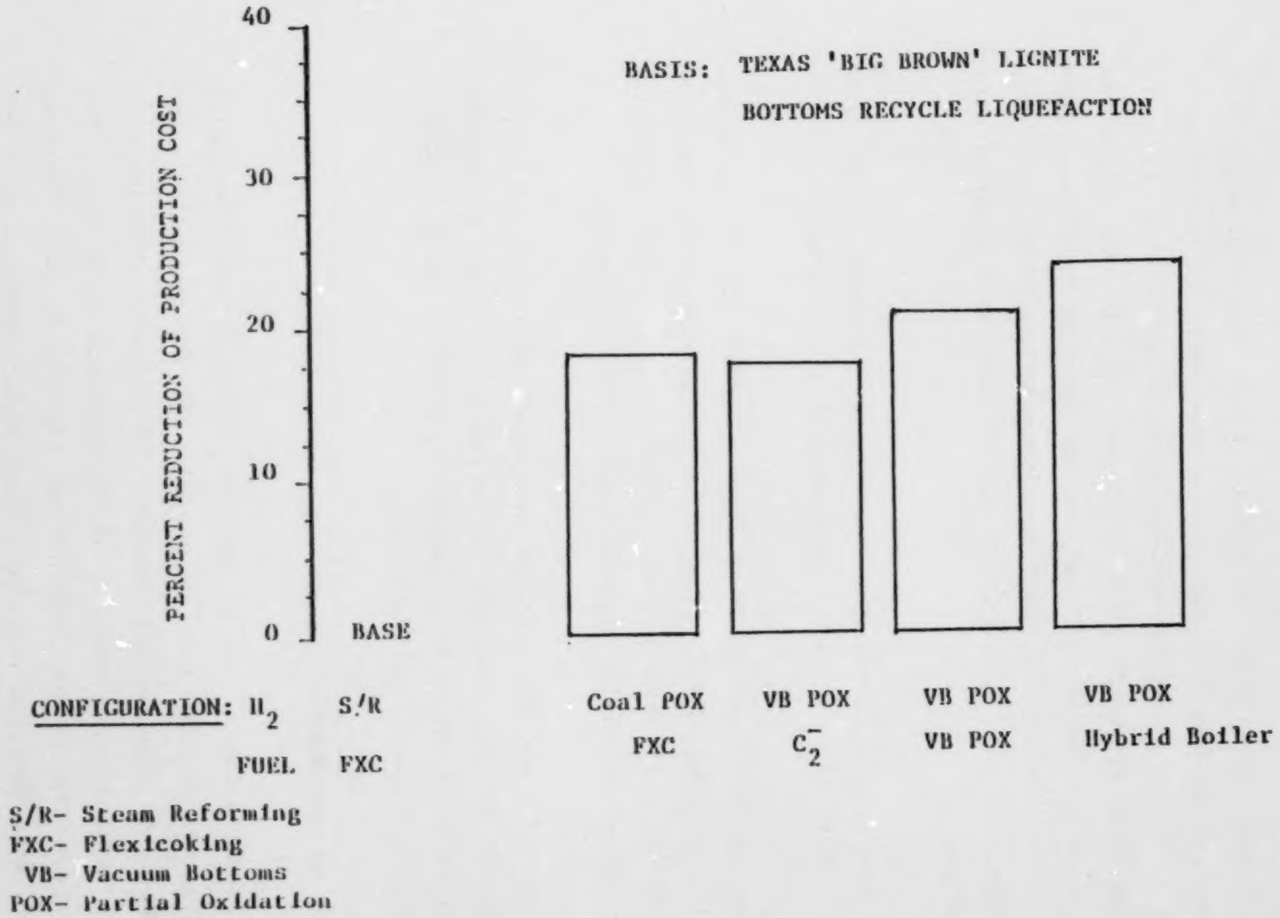

Figure 2-13 Evaluation of Process Alternatives to Meet Hydrogen and Fuel Requirements

Source: Analysis of Alternative Processes for Manufacture of Fuel and Hydrogen in a Coal Liquefaction Plant - The EDS Process, Exxon Research and Engineering Co.

2.4.2.4 Products Yields, Distribution and Efficiencies

Typical EDS liquid yields of C_5-525°C material for different rank of coal, as estimated by the process developer, are presented in Table 2-20 for a once-through liquefaction mode. Typical analyses of these coals are given in Appendix B. In this configuration about 2.6 bbl/t MAF coal can be produced from Texas lignite. The yield is 11% lower when compared with the yield from bituminous Illinois No.6 coal or only 5% lower when compared with Wyodak coal.

On MF basis, the decrease in total liquid product yields from Texas lignite is 20 wt% when compared with Illinois coal.

However, in the case of Texas lignite, about 45% of the total liquid is naphtha (C_5-200°C) which is 13% more when compared with Illinois No.6 bituminous coal. This indicates that lignites having higher reactivity produces more lighter products.

The "overall" and the "liquid products" efficiency of Texas lignite in the EDS process, as estimated by the process developer,is 58.3% and 52.6% respectively, which is about 2% higher compared to Illinois coal efficiency.

- **EDS Bottoms Recycle Liquefaction Configuration**

As indicated earlier, one of the EDS process alternatives being investigated is the partial recycle of vacuum distillation bottoms to the liquefaction reactor. In this configuration, the overall conversion of coal to liquid products is higher. The bench-scale experiments with Texas lignite in the "bottoms recycle" mode indicate that the liquid yields increase by 8 wt%

Table 2-20 Comparative Product Yield and Efficiencies for Different Rank of Coals in the EDS Process

	Bituminious Illinois No.6	Subbituminious Wyodak	Lignite Texas Big Brown
Coal (as received)			
Moisture, wt%	14.0	29.0	28.4
Ash, wt% MF coal	8.9	7.8	16.4
Operating Conditions			
Pressure, MPa	13.8	13.8	13.8
Residence Time, min	25.0	40.0	60.0
Hydrogen Treat Rate, wt% MAF	4.0	4.0	4.0
Solvent-to-Coal Ratio	1.6	1.2	1.0
Liquid Product Yield			
bbl/t as-received coal	2.3	1.9	1.7
bbl/t MF coal	2.7	2.5	2.0
bbl/t MAF coal	3.0	2.8	2.6
Naphtha in Total Liquid, wt%	32.3	39.1	44.7
Thermal Efficiency, %			
Overall	56.0	59.6	58.3
Liquid Products	51.0	52.5	52.6

Source: DOE Report FE-2893-35, September 1979 on MF coal basis (or 8.5 wt% MAF coal) when 50% of the vacuum bottoms are recycled to the reactor. However, this increase is achieved on account of higher pressure (13.8 MPa vs 10.3 MPa), increased residence time (100 minutes vs 60 minutes) and increased hydrogen consumption (7.2 wt% vs 4.8 wt% MAF coal).For this configuration, the liquid product yields for a conceptual, self-sufficient plant are not available yet. Typical reactor yields of liquid products for different rank coals are presented in Table 2-21. For comparison purposes, reactor yields in the "once-through liquefaction" mode are also included in this table.

In the "bottoms recycle" mode, the liquid yield for Texas lignite is about 5% lower on MAF basis as compared to bituminous coal. The bottoms-recycle liquefaction mode, however, is also associated with increases in the capital cost. Data on capital cost increases for Texas lignite are not available. For a solvent/coal/bottoms ratio of 2.5/1/1 with Illinois No.6 coal, an increase of about 62% in the capital cost was estimated by the Exxon Research and Engineering Company (DOE report FE-2893-49, July 1980). About one-half of the increase is due to the higher operating pressure in the liquefaction reactor. The balance is due to a larger amount of material to be handled.

The overall material balance for a conceptual, commercial, integrated, self-sufficient EDS plant operating with Estevan lignite (without bottoms recycle) is presented in Table 2-22. These results are preliminary and do not necessarily reflect optimal commercial plant setting conditions. The basis used is 25,000 t/d of as-received lignite to the plant.

Table 2-21 Comparison of "Once-Through" and "Bottoms-Recycle" Liquefaction Reactor Yields in the EDS Process

	Process Conditions*	EDS Run No.	Yields, wt% MAF Coal**		
			C_4-200°C	200-525°C	Total
Once-Through Liquefaction					
Bituminous (Illinois No.6)	10.3, 40, 4.4, 1.2/1/0	256-264	20.0	16.4	36.4
Subbituminous (Wyodak)	10.3, 60, 4.4, 1.6/1/0	267-270	24.6	7.5	32.1
Lignite (Texas Big Brown)	10.3, 60, 4.8, 1.6/1/0	344	25.7	12.4	38.1
Bottoms-Recycle Liquefaction					
Bituminous (Illinois No.6)	13.8, 60, 8.8, 1.8/1/1	308-311	34.8	14.2	49.0
Subbituminous (Wyodak)	13.8, 100, 8.8, 2.5/1/1	312-315	30.6	13.7	44.3
Lignite (Texas Big Brown)	13.8, 100, 7.2, 1.6/1/0.5	338-340	28.1	18.5	46.6

* Conditions refer to pressure (MPa), nominal residence time (min), hydrogen treat rate (wt% MAF coal) and solvent/coal/bottoms ratio.

** Product yields reported by EDS on MF basis have been converted to MAF basis considering 9.1 wt% ash MF coal for Illinois No. 6 coal, 9.71 wt% ash for Wyodak coal, and 16.4 wt% ash for Texas Big Brown lignite.

The utility requirements from outside sources are water and electricity. The amount of raw water and power required are about 360 L/s and 110 MW respectively. About 4 wt% (dry basis) of lignite feed is required to raise steam in offsite coal-burning boilers. The entire process fuel requirements are met by the low-CV fuel gas produced in the FLEXICOKER.

The overall thermal efficiency of the EDS process for Estevan lignite is 53.4% based on the overall product yield given in Table 2-22. Similarly, the liquid product efficiency is 48.2%.

2.4.2.5 Suitability/Availability of Process Inputs

Comparison of the analyses of Saskatchewan Estevan and Texas Big Brown lignites shows that the two lignites are quite similar. With respect to the elemental composition, Estevan lignite has similar carbon, hydrogen and nitrogen contents. The sulfur content is 50% less and the oxygen content is 20% more when compared with Texas lignite. Concerning the proximate analysis, Estevan lignite contains about 5% less ash compared to Texas lignite. Of the ash elements, Estevan contains 15% more SiO_2 and 13% less CaO content. Petrographically, Estevan lignite contains 8 vol% more reactive macerals.

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* Analyses of Estevan and Willow Bunch lignites, see Appendix A; for Texas Big Brown, see Appendix B.

Table 2-22 Estimated Product Yield and Efficiencies in the EDS Plant
(Basis: 25,000 t/d "as-received" coal)

	Estevan	Texas Lignite Big Brown
Coal (as received)		
Moisture, wt%	35.0	28.4
Ash, wt% of MF coal	15.4	16.4
Coal Fed to Reactor		
Moisture, wt%	4.0	4.0
Ash, wt% of MF coal	15.4	16.4
Input		
MF Coal to Reactor, t/d	15,625	17,212
MF Coal to Gasifiers, t/d	-	-
MF Coal to Offsites, t/d	625	688
Electricity, MW	110	110
Output, t/d (bbl/t)		
Propane	123	135
Butane	199	219
Naphtha (C_5-200°C)	2,210 (17,365)	2,435 (19,133)
Fuel Oil (200-525°C)	2,746 (16,782)	3,025 (18,487)
Sulfur	48	200
Phenols	182	53
Ammonia	114	126
Liquid Product Yield (C_5-525°C)		
bbl/t as-received coal to plant	1.19	1.31
bbl/t MF coal to plant	1.83	1.83
bbl/t MAF coal to plant	2.16	2.16
Naphtha in Total liquid, wt%	44.7	44.7
Thermal Efficiency, %		
Overall	53.4	58.3
Liquid Products	48.2	52.6

Notes:
1. The data presented are preliminary estimates based on ER&E data utilized for the process variables screening study using the PAM computer model, Case No. 864.
2. Liquefaction conditions: 449°C, 25 minutes nominal residence time, once-through process configuration including application of FLEXICOKING, and H_2 production by steam reforming.
3. Lignite to offsite estimated on the basis of Conceptual Commercial Plant Design developed by ER&E for Illinois No.6 Coal
4. Plant fuel requirements are met by low-CV gas from FLEXICOKING.
5. Product heating values are approximated average values.
6. Electricity purchased from outside source, 1 kW = 10,000 Btu/h (10,550 kJ/h) is assumed as conversion factor.
7. Fuel oil equivalent factor of 6MM Btu/bbl (6.33 GJ/bbl)has been assumed.

- **Moisture Reduction**

Run-of-mine coal looses its moisture very rapidly down to equilibrium moisture on exposure to air because much of the water is on the surface. Further reduction of the moisture content in lignite, when air-drying is used, could result in shrinkage and collapse of capillaries due to the decrease in the vapor pressure of water within the capillaries. As a result, the pore structure is changed and the liquefaction reactivity of the lignite is reduced. Also, during conventional air-drying of lignite, highly reactive sites are exposed to oxidation which further reduces the reactivity.

To avoid the collapse of capillaries and surface oxidation and to preserve the reactive surface areas, a coal/oil slurry drying method is employed in the EDS process which appears to circumvent the problems of air drying. The coal is dried simultaneously while preparing the coal/oil slurry feed for the liquefaction reactor. In this system, crushed coal and hot recycle solvent are fed to a well-stired tank maintained at about 121-178°C and near 101 kPa pressure. Moisture evolves from the coal particles and bubbles up through the hot solvent. No oxygen comes in contact with the hot coal. The evolving moisture generates a foam. So, space is provided for this. The evolved moisture along with some stripped solvent vapor is withdrawn from the slurry drier and condensed. Condensed solvent is returned to the drier. Adequate residence time is provided to dry the coal slurry to less than 4 wt% on dry coal feed. The heat required to vaporize the water from the coal is provided by recycling a portion of the slurry through a heat exchanger.

- **Operability with Lignites**

In the EDS process, two major operational concerns are related to the processing of low-rank coals. The first concern is the vacuum-bottoms viscosity and the second relates to the calcium content of feed coal which causes reactor wall deposits and line pluggings.

- Vacuum Bottoms Viscosities

The vacuum bottoms viscosity is generally related to the oxygen content of the bottoms and, therefore, to the liquefaction severity; with lower viscosities obtained at higher severities. Below a 40-minute liquefaction residence time, the bottoms viscosity for Texas Big Brown lignite is about 100 poise (10 kg/m/s) which is in the region considered difficult for pumping equipment. Beyond the 40-minute residence time, however, the viscosity of the vacuum bottoms begins to decrease (see Figure 2-14). Recycling a portion of the vacuum bottoms to the liquefaction reactor has also been found to reduce the bottoms viscosity. This is attributed to a reduction in the concentration of polar groups (or oxygen functionalities) present in the vacuum bottoms. Without bottoms recycling, for Saskatchewan lignites in the EDS process, a residence time of around 60 minutes in the liquefaction reactor would be required. Alternatively, bottoms recycle liquefaction would be necessary if the residence time is kept between 25 to 60 minutes.

- Calcium Carbonate Deposition

Past EDS experience with Wyodak coal, Texas and North Dakota lignites has indicated the formation of $CaCO_3$ wall-scale and free-flowing aggregates called "oolites" in the EDS reactor. The amount of $CaCO_3$ formed is directly related to the calcium content initially present in the feed coal as shown in Figure 2-15. Total solids and oolite accumulation rates for Texas lignite are higher than those measured for Wyodak coal and North Dakota lignite. This is consistent with the elemental composition of these coals. The calcium content of Estevan lignite (1.28 wt% MF coal) is almost the same as Wyodak coal and the Ca content of Willow Bunch Lignite (2.15 wt% MF coal) is similar to Texas Big Brown coal. The high level of Ca content in Saskatchewan lignites may cause operational problems. One method that has been found effective in eliminating the $CaCO_3$ problem is to pretreat the coal with gaseous sulfur dioxide. Saskatchewan lignites would probably require such pretreament in the EDS process.

- **Utility Requirements**

In terms of process inputs, the EDS conceptual commercial plant operating with Estevan or Willow Bunch lignite would require about 110 MW electricity and 360 L/s of raw water from outside source. Requirement of utilities in these amounts is not expected to be a problem.

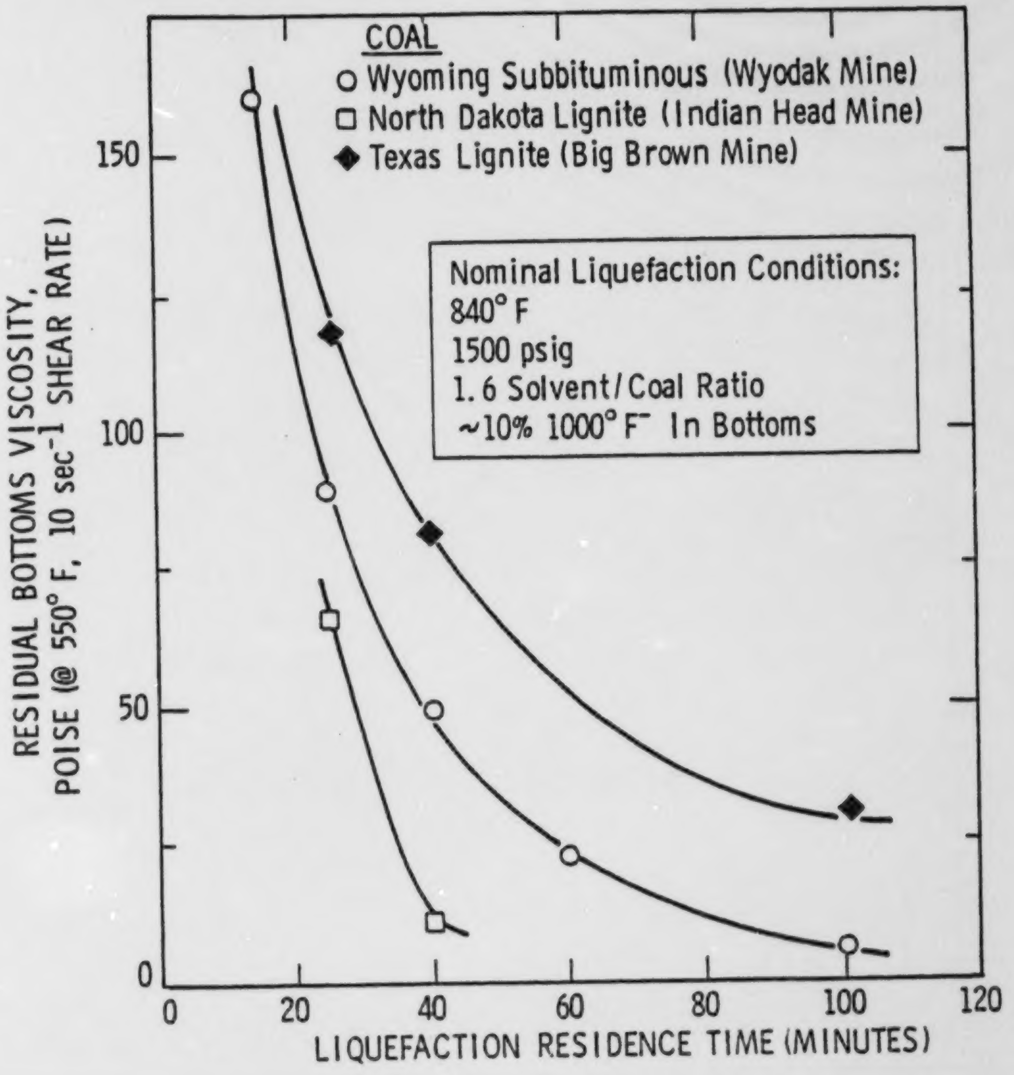

Figure 2-14 Effect of Liquefaction Residence Time on Residual Bottoms Viscosity

Source: DOE Report FE-2893-21, December 1978

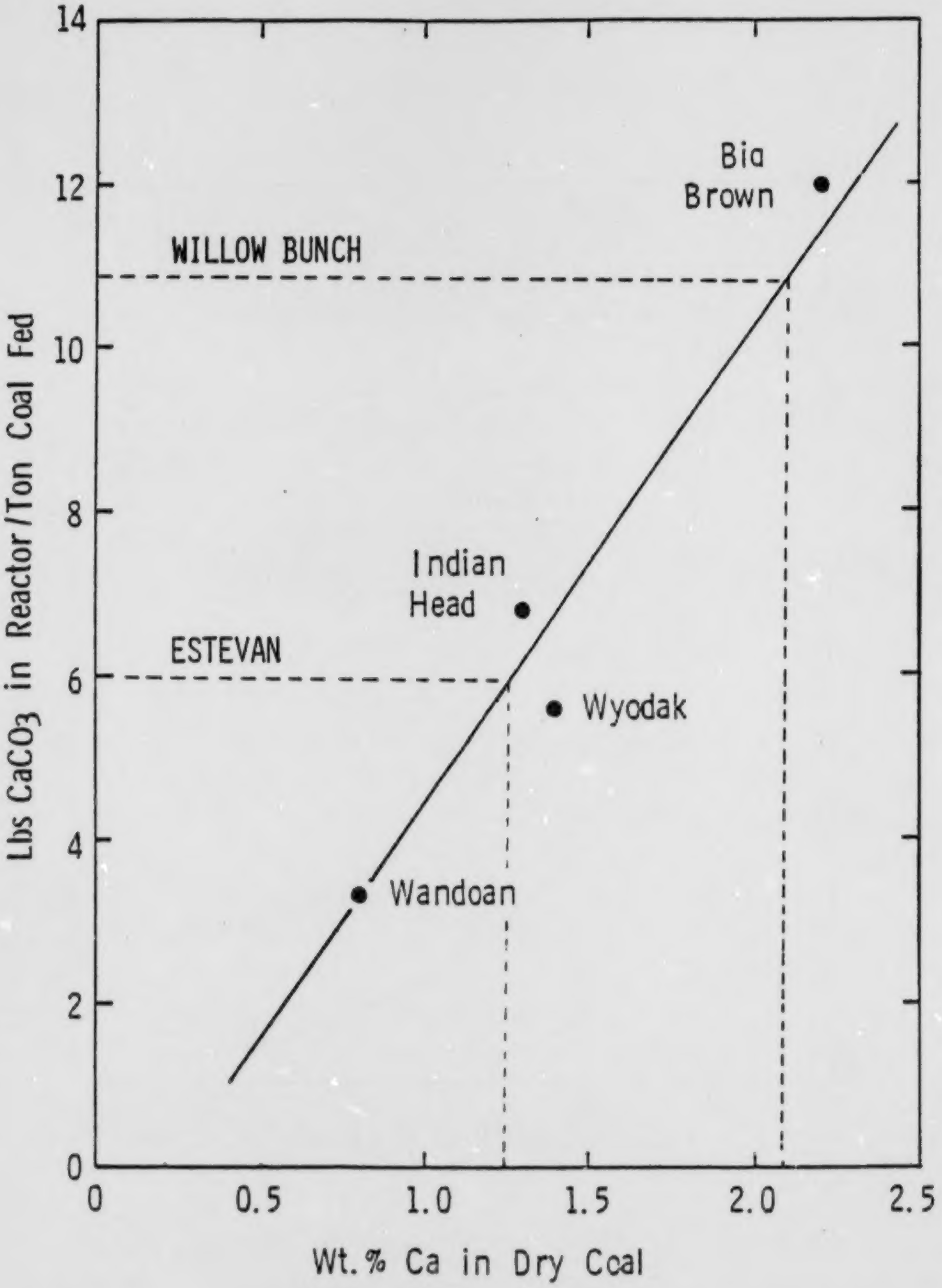

Figure 2-15 Calcium Carbonate Deposition From Lower Rank Coals In The EDS Liquefaction Reactor

Source: DOE Report FE-2893-35, September 1979

2.4.2.6 Product Quality, Utilization and Upgrading Schemes

The product quality, utilization and upgrading schemes for EDS naphtha, mid-distillate and vacuum gas-oil (distillate fuel oil) from bituminous and subbituminous coals have been established through extensive testing and characterization. Similar studies for products obtained from lignites have yet to be carried out.

Table 2-23 presents the API gravity, TBP distillation and elemental composition of naphtha (C_5-200°C), mid-distillate (200-370°C) and vacuum gas-oil (370- 525°C) produced from Wyodak coal in the EDS process. Similar analyses data for products from Texas lignite, except for vacuum gas-oil, are also presented in this table. The elemental analysis for naphtha from lignite reveals much higher levels of sulfur, nitrogen and oxygen ascompared to Wyodak naphtha. This is consistent with the heteroatoms present in the feed coal. Since naphtha from lignite may contain higher levels of heteroatoms than Wyodak naphtha, hydrotreating severity would be very high (71 to 85 m^3/bbl H_2 consumption).

The mid-distillate stream represents product recovered downstream of the solvent hydrotreater (as excess hydrotreated solvent) in 1-t/d pilot plant. Typically this stream contains about 75 wt% aromatics and 25 wt% saturates. Of the 25 wt% saturates, the paraffin content is only 4 wt%. This product stream can be directly used as No.2 fuel oil (domestic heating oil) or as a stationary turbine fuel. However, because EDS mid-distillate does not meet the conventional No. 2 fuel oil gravity specification (30° API), reduction in the aromatics content from about

75 wt% in the original product to about 20 wt% is required. This can be achieved through conventional hydroprocessing at a severity level corresponding to about 71 m^3/bbl H_2 consumption.

Alternatively, mid-distillate can be extracted with an appropriate solvent to separate a saturate-rich raffinate containing about 20 wt% aromatics. The raffinate can be used as a jet-fuel. The aromatic-rich extract (10 wt% non-aromatics) may be used as fuel-oil blendstock or can be recycled to the liquefaction reactor.

The distillate fuel oil (vacuum gas-oil) requires severe hydrotreatment to reduce the nitrogen and sulfur content to a level suitable for use as a stationary combustion fuel. Typically a high hydrotreating severity (65 m^3/bbl H_2) is required to achieve over 60% denitrogenation and 90% desulfurization. The vacuum gas-oil (VGO) stream, as indicated above, requires deep denitrogenation to produce acceptable fuel oil. Deep denitrogenation is achieved at the expense of H_2 consumption. The severe hydrotreating conditions not only provide denitrogenation but also an appreciable level of conversion to lighter products. For these reasons, it may be advantageous to combine the mid-distillate and distillate fuel oil products and then hydrotreat the combined stream.

Some quantitative hydrotreatment data for the combined product stream from Illinois coal are as follows: the catalytic cracking (zero H_2 consumption) results in a 24 LV% yield of 100 RON motor gasoline blendstock. The yield of this blendstock can be increased to 54 LV% through mild hydrotreating (45 m^3/bbl H_2 consumption). The severe hydrotreatment (65 m^3/bbl H_2

Table 2-23 Properties of EDS Product Streams

	Wyodak Coal			Texas Big Brown Lignite		
	Naphtha C_5-200°C	Mid-distillate 200-370°C	Vacuum Gas Oil 370-525°C	Naphtha C_5-200°C	Mid-Distillate 200-370°C	Vacuum Gas Oil 370-525°C
Gravity, °API	42.1	16.1	-5.3	45.6	14.4	N.A.
Distillation, °C						
5%	43	185	318	-	202	N.A.
50%	141	238	399	146	-	
95%	188	377	516	195	392	
Elemental Analysis, wt%						
Carbon	86.0	89.6	89.3	83.01	89.9	N.A.
Hydrogen	12.6	9.5	7.5	12.6	9.6	
Oxygen	2.7	1.14	2.45	3.93	0.50	
Nitrogen	0.16	0.01	0.05	0.30	0.01	
Sulfur	0.05	0.01	0.05	0.17	10 ppm	

Notes: 1. Products obtained from the 1-t/d pilot plant for both coals

2. The liquefaction conditions are:

449°C, 60 minutes, 10.3 MPa, 1.6 solvent/coal, 4 wt% H_2 for Wyodak coal
449°C, 60 minutes, 13.8 MPa, 1.6 solvent/coal, 4 wt% H_2 for Texas lignite

consumption) increases the yield to 85 LV% of feed. A hydrocracking operation selective to motor gasoline production indicates the feasibility of achieving nearly 130 LV% of 100 RON blendstock at an H_2 consumption level of 99 m^3/bbl.

2.4.2.7 Areas Requiring Further Development

ER&E has operated the 227-t/d pilot plant with bituminous Illinois No.6 coal for about 12 months with a longest continuous run of 60 days. During this operating experience several problems were encountered particularly in the areas of slurry drying and preheating.

The slurry drying horizontal heat exchangers used for providing heat to the slurry drier, via a circulation loop, were often plugged. The plugging was rapid during the initial start-up. Several attempts to modify these horizontal exchangers failed. At present, redesigned heat exchangers having a four-pass vertical configuration are being tested.

The coking of the slurry in the preheat furnace has been a significant problem. Improper distribution of coal, solvent and gas resulted in a higher viscosity, a lower heat-transfer coefficient and a higher film-temperature which are believed to be the cause of coking in the preheat furnace tubes.

The reactor control and performance particularly with lignitic coals are another area of concern. The problems are due to higher oxygen and calcium contents in these coals. As has been observed, a higher oxygen contents in the feed coal results in a higher viscosity of the vacuum bottoms. The high calcium content results in the formation and deposition of significant amounts of calcium carbonate inside the reactor and in downstream equipments. Mechanical systems for solid removal have been found to alleviate the deposit problem in the PDU operations. These systems are presently being tested in the pilot plant with Wyodak coal.

2.4.2.8 Status and Commercialization Prospects

The EDS process is being developed as a joint undertaking between the United States Government and a consortium of private industries. These development project sponsors are: The U.S.Department of Energy, Exxon Company, U.S.A., Electric Power Research Institute (EPRI), the Japan Coal Liquefaction Development Company (JCLD) (a consortium of twelve Japanese companies), Phillips Coal Company, ARCO Coal Company, Ruhrkohle AG, (a coal producer from the Federal Republic of Germany), and Agip S.p.A. (a subsidiary of the Italian national oil company).

The principal objective of the EDS project is to obtain data to allow a commercial-scale design of the EDS process by 1983. This is being achieved through an integrated process development program involving pre-development work at bench-scale level, development work on a 1-t/d pilot plant and operation of a 227-t/d pilot unit, the Exxon Coal Liquefaction Plant (ECLP). The pilot plant has been operated for 11 months with Illinois No. 6 coal.

The plant was restarted to evaluate the suitability of Wyoming subbituminous coal, including the demonstration of the "bottoms-recycling" mode. When the Wyoming coal run is finished, the plant will be converted for Texas lignite processing.

The 227-t/d coal feedrate and associated solvent and gas feed rates for the ECLP were selected to achieve liquid and gas velocities close to those assumed for a commercial EDS reactor. These rates were determined using pre-developed correlations. Once their validity is confirmed, or modified as a result of the ECLP test program, these correlations would then be used to size the reactors for a commercial plant at a 30-times higher coal feed rate than in ECLP, i.e. 6805 t/d coal feed per reactor train.

The coal slurry at ECLP is heated to the reaction temperature in a vertical tube furnace, which has a similar design to that used in a petroleum refinery. There are indications that the slurry will coke, and foul the inside surface of the furnace tubes, thereby limiting the plant operability. One objective of the ECLP operation is to demonstrate that such a commercial-type furnace can be applied to the preheating of coal slurries.

Several different types of pumps and a variety of operating conditions were specified for ECLP. The objective is to obtain pump design data for coal liquids such as net positive suction head and to specify the types of pumps for different services in a coal liquefaction plant. Materials of construction can have a significant effect on the service factor and cost of a commercial plant. The materials selected for ECLP are those used in a commercial plant, rather than the more exotic materials which may be used to ensure a high service factor in a pilot unit. Lastly, line sizes and flow velocities are also in the range of an expected commercial operation. The objective is to confirm the adequacy particle saltation correlations and to gather erosion data on the commercial process and fluid dynamic conditions.

The outlook of the ECLP operation as of April 1, 1981 was positive. Concerning the materials of construction, a few localized corrosion problems were observed in the pilot plant equipment. Isolated instances of severe erosion experienced during shakedown appear to have been ameliorated by the installation of fiber-reinforced refractory linings. The let-down valves, which reduce the slurry pressure from 13.8 MPa to about atmospheric pressure in one step, have exhibited an excellent performance with only minor erosion noted after more than 100 days of service. Specially selected high-pressure block valves and check valves have performed adequately. Data required to design environmental control systems for a commercial-scale plant, including air, water, and noise control systems have been obtained.

The complete analysis of all ECLP tests and operational data gathered during its operation will require about six months elapsed time upon termination of the data-taking work.

Recently, the sponsors of the EDS process have decided to terminate the development of the $64-million FLEXICOKING program. Under the new program, ER&E is evaluating vacuum-bottoms recycling and partial oxidation of vacuum bottoms in Texaco or Shell gasifiers. The hybrid-boiler concept, in which the bottoms are fed to a boiler to raise both steam and to preheat the coal slurry before it is fed to the liquefaction reactor, is also being investigated.

Current projections based on the ECLP operation and the schedule for the bottoms processing option indicate that a basis for a commercial plant design may be available in the first quarter of 1983 assuming successful continuation of the program.

SNC

2.5 LIQUID SOLVENT EXTRACTION PROCESS

Two direct catalytic coal liquefaction processes are being developed in England by the National Coal Board (NCB) at the Coal Research Establishment,Stoke Orchard, Cheltenham since 1973; namely the Supercritical Gas Extraction (SGE) and Liquid Solvent Extraction (LSE) processes. Since the SGE process is beyond the scope of this study, only the technical assessment of the LSE process is presented.

In the Liquid Solvent Extraction (LSE) process, pulverized coal is mixed with recycle distillate oil, preheated, and fed into a coal extraction reactor operating at about 400°C and 2 MPa. The reactor effluents are condensed and separated into gases/vapors and solid/liquid slurry streams by conventional flash separation methods. The solid residues (unreacted coal, char, ash) from the slurry are removed by filtration and gasified for the production of make-up hydrogen for the process. The filtrate (extract) is catalytically hydrocracked and then fractionated to yield naphtha (C_5-200°C), middle oil (200-300°C) and heavy oil ($>$300°C). The naphtha and middle oil may be further upgraded to gasoline and diesel fuel, and the heavy oil is recycled as a solvent for the process. The coproducts are SNG and LPG. Byproducts are sulfur, ammonia and crude phenols.

The LSE process is essentially a two-stage process comprising an "extraction" stage where coal extract is separated from mineral matter and ash, and a subsequent "extract hydrogenation" stage where the coal-derived liquids are stabilized before fractionation. This approach has the advantage that for a given coal feedstock it is possible to optimize the conditions in each stage so as to produce the desired product yield with minimum hydrogen consumption.

2.5 LIQUID SOLVENT EXTRACTION PROCESS (cont'd)

Other main features of the process are:

- The extraction reactor operates at a relatively low temperature and pressure (400°C and 2 MPa), hence capital and maintenance costs are lower.

- Separation of solids (unreacted coal, char, ash) by filtration allows maximum recovery of coal-derived liquid.

- The absence of solids in the coal-derived liquid permits the use of more selective catalysts in the hydrogenation stage, since they are not liable to be deactivated or poisoned by mineral matter.

2.5.1 History and Development of the Process

Research and development work on the extraction of coal with anthracene oil and other tar oils began in the mid-60s at the British Coal Utilization Research Association (now the Coal Utilization Research Laboratories of the National Coal Board) to produce smokeless fuels for burning purposes. Alternatively, the manufacture of carbon products from the extract, particularly electrode coke for steel-smelting and aluminium-smelting furnaces, was also examined. Extraction experiments and investigations were initially carried out in a laboratory-scale unit (5 kg/h of coal) and later in a 30-kg/h bench-scale extraction unit constructed at the Coal Research Establishment (CRE).

Following the oil crisis of 1973/74, efforts were directed towards production of transportation fuels and/or petrochemical feedstocks. Therefore, a second-stage catalytic hydrocracking step of the coal-derived extract (instead of coking) and refining of products were added to the extraction stage. A separate catalytic hydrocracking unit, with a throughput of 1 kg/h of extract was constructed at CRE with financial assistance from British Petroleum Company. More than 50 hydrocracking catalysts have been evaluated at CRE so far. Upgrading and refining schemes of the hydrocrackate (syncrude) to gasoline, diesel fuel and jet fuel have also been investigated.

Based on the bench-scale data, a 22.7-t/d pilot plant to be built at Point-of-Ayr, North Wales, has been proposed. The detailed process design and cost estimation have been carried out.

2.5.1.1 Bench-Scale and PDU Program

The LSE process has been tested on a laboratory scale (5 kg/h of coal) and in a bench-scale experimental unit (30 kg/h of coal). The basic process engineering of the bench-scale unit was done by the Coal Research Establishment of the NCB, while the detailed engineering and construction was carried out by Woodall-Duckham, a subsidiary of Babcock and Wilcox, United Kingdom. The mechanical construction and commissioning of this continuous extraction facility was completed in September 1977 at a total cost of about 1.5 million dollars.

Figure 2-16 is a schematic diagram of the bench-scale unit of the LSE process. Coal, pulverized to a nominal size of 0.21 mm (210 microns) and dried to 5 wt% moisture, is mixed with recycle solvent in the oil-to-coal ratio of 2 to 4. The resulting slurry is preheated to 400°C and fed into the extraction reactor.

The reactor effluents are filtered to remove the solid residues (ash, unconverted coal and char). A portion of the filtrate (solvent and coal-derived extract) is sent for solvent recovery and the remaining portion is fed into a two-stage catalytic hydrocracker. The solvent-free coal-derived extract from the solvent evaporator is also sent to the hydrocracker. The hydrogenated extract is fractionated into light oil (C_5-200°C), middle oil (200-300°C) and heavy oil ($>$300°C) which is recycled to the reactor.

To date, extensive experiments have been carried out to investigate the influence of operating conditions on extract yield and quality. Coal residence time, temperature and pressure are the most significant variables influencing the extract yield. In contrast, a five-fold increase in the solvent flowrate has only a small effect ($\simeq$ 10%) on the extract yield and a similar change in particle size is even less effective.

The type of solvent used affects not only the yield but also the quality of the extract. Initial experiments were carried out with anthracene oil and other similar petroleum-derived fractions with moderate success (50% extraction of MAF coal). Continued experiments revealed that partially hydrogenated coal-derived heavy oil was better for three main reasons:

- It has a low volatility; at the high-temperature and low-pressure operating conditions in the extraction reactor, the solvent loss is low.

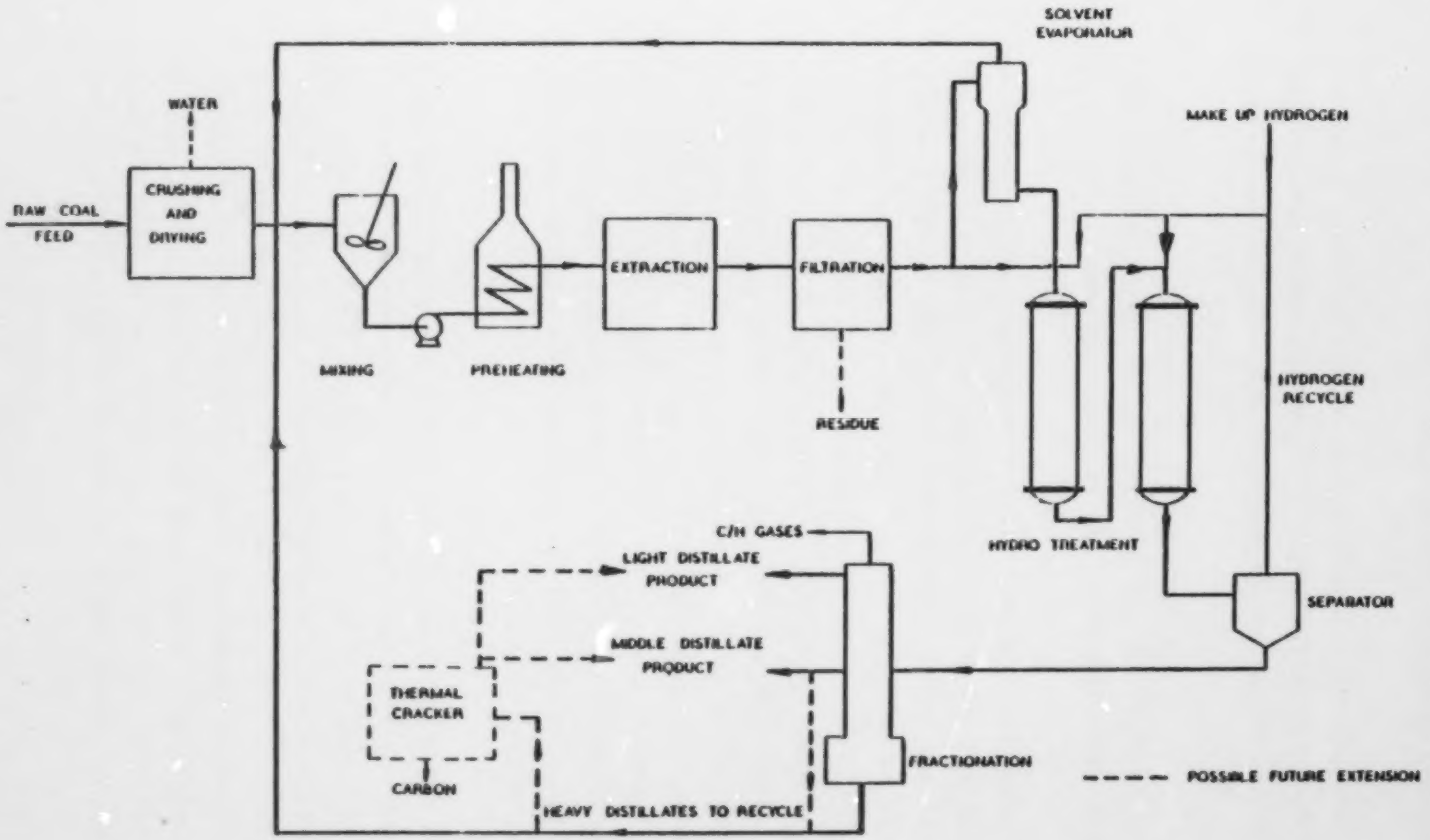

Figure 2-16 Schematic Diagram of the LSE Bench-Scale Unit (30 kg/h of coal)

Source: National Coal Board, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", Feb. 1981.

- The hydrogen-donor effect of the solvent produces an extract of lower viscosity and molecular weight, consequently higher filtration rates can be achieved and more lighter products obtained.

- A higher extraction (up to 95 wt% of MAF coal) and net liquid product yield (50 wt% of MAF coal) are achieved.

Different techniques of solid/liquid separation such as hydroclones, filtration and solvent precipation have been evaluated. Of these available methods, NCB prefers the filtration route because fundamental investigations since 1976 have allowed optimization of the extraction-reactor operating conditions to give filtration rates that should make this approach viable on a large scale. Moreover, recent developments in the hot-filtration technology are leading to new filters, specially designed for coal slurry application. Although these filters are expensive (10% of the total processing cost), NCB believes that this method of solid/liquid separation is cost-effective and has a potential for cost savings. At present, testing has been carried out using 30-kg/h prototype filters.

As part of the bench-scale process development program, extensive analyses and characterizations of the primary extraction products (extract and residue) have been carried out. Several tests have been made to evaluate the potential of these products for further processing, i.e., hydrocracking of coal extract and gasification and combustion of residues. Catalytic hydrocracking of extract has been performed in a 2-L fixed-bed hydrocracker. Gasification tests on residual char in steam, air and carbon dioxide have been carried out in 0.04-m and 0.15-m fluidized-bed units confirming the potential of these residues as gasification feedstocks. The suitability of the residues for combustion has been examined by carrying out ignition and burn-out tests in 0.15-m fluidized-bed combustor.

2.5.1.2 Pilot-Plant Program

The National Coal Board with financial assistance from U.K. Department of Energy, European Coal and Steel Community, Phillips Oil Co., and British Petroleum is engaged in a pilot-plant program since 1979. Two options were under consideration, i.e. to build a combined pilot plant to investigate both the LSE and SGE processes or to investigate one of the two process alternatives. Recently, a decision in this respect was made in favor of the LSE process only.

The LSE pilot plant will have a capacity of processing 22.7 t/d of coal. The proposed location of the plant is a 10.5 hectare site near the Point-of-Ayr collieries in North Wales. The process design and detailed cost estimation have been carried out by Mathews-Hall Ortech Ltd.

Figure 2-17 is a process flow diagram of the proposed LSE pilot plant. Raw coal from storage bunkers is mechanically conveyed to a pulverizer/dryer where the coal is crushed and dried to 5 wt% moisture. The dried and pulverized coal is mixed with the recycle solvent in the slurry preparation vessel. The slurry containing 30 to 40 wt% solids is pumped to the reactor pressure ($\simeq$ 2 MPa), preheated to 400°C and fed into a well-stirred extraction reactor. The reactor effluent, after flashing to remove light hydrocarbon gases, is sent to leaf-type filtration units for removal of solid residues. The residues (filter cake) are dried and then sent to an incinerator for disposal. In a commercial plant, the residues will be gasified to produce make-up hydrogen for the process.

The filtrate (extract and solvent) is sent to the solvent recovery section. A small portion of the recovered solvent is used for back-washing of the filter and the remainder is sent to the slurry preparation unit. The extract is pumped to a catalytic hydrocracker operating at 420°C and 20 MPa.

The hydrogenated extract is preheated and fed into a fractionator to produce distillate products. The naphtha is stabilized and the middle oil is steam-stripped before storage. Some oil is also recovered from the fractionator and used as a wash solvent for the filters. The fractionator bottoms are sent to a vacuum distillation tower to recover solvent for recycle. The residual oil (>525°C) from the vacuum tower is sent to a delayed coker to recover some distillates and to produce coke.

2.5.2 **Technical Review**

2.5.2.1 Types of Coals Investigated

Different types and ranks of coals have been tested in the LSE process, however, the main effort has been directed towards the evaluation of high-volatile, non-caking butuminous U.K. coals. Some preliminary investigations have been done on coals from other countries such as U.S.A., Australia, Canada and Turkey

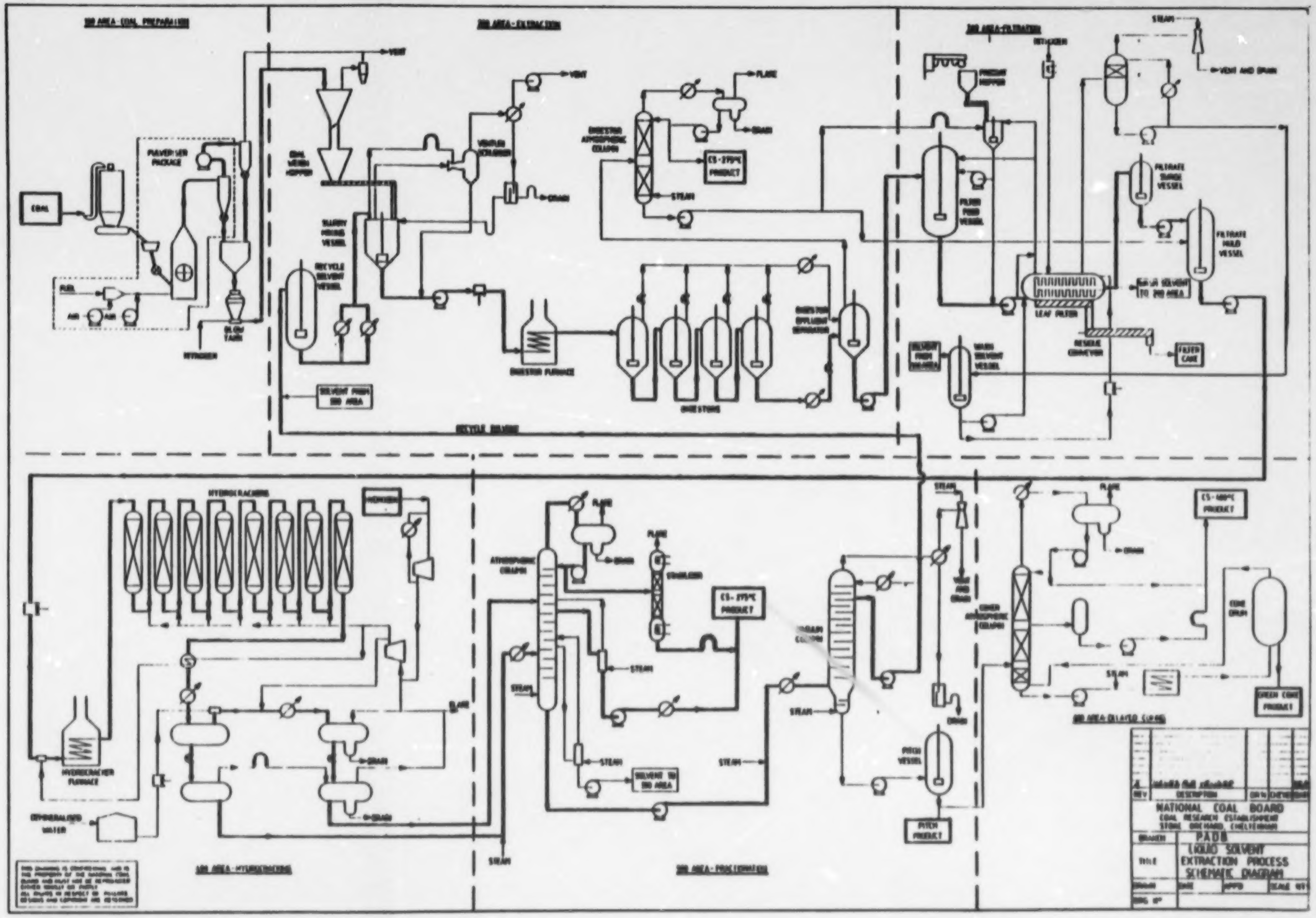

Figure 2-17 Process Flow Diagram of the LSE Proposed Pilot Plant

Source: W.P. Syrett, NCB, Communication to J.A. Mikhlin, SNC Inc, NCB Ref. No. 496/2, 7/3 Jan. 25, 1982

Typical analyses of a U.K. bituminous coal (Annesley, NCB Coal-Rank-Code 702) and an Australian brown coal are given in Appendix B.

The U.K. bituminous coal (CRC 702) has 10 wt% moisture on as-received basis and about 11 wt% ash on MF basis. A detailed technical and economic evaluation of the LSE process carried out by NCB was based on the utilization of this coal.

2.5.2.2 Overall Process Configuration of a Conceptual, Commercial Plant for the LSE Process

Figure 2-18 is a simplified flow diagram of the LSE process. The conceptual design of the plant is based on the utilization of a high-volatile bituminous coal (CRC 702). Differences in the processing scheme for Saskatchewan lignites as suggested by NCB* are also presented below.

- **Coal Preparation**

The coal preparation section of the plant includes facilities for receiving coal, storage, crushing and drying of the coal to desired specifications. In the case of Saskatchewan lignite, it is ground to minus 0.075-mm size and dried to 15 wt%. Coal for steam and power generation is dried to 5 wt%.

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* W.P. Syrett, NCB, Personal Communication to J.A. Mikhlin, SNC Inc., NCB Ref. No. DR 3000, DR 496/2, October 1981

- **Coal Liquefaction and Product Recovery**

The dried and pulverized coal is mixed with recycle oil in an oil-to-coal ratio of 3. For Saskatchewan lignites, an oil-to-coal ratio of 2, as suggested by the process developers, was considered. In a mixing tank, the residual moisture present in the feed coal is removed by a slurry-drying method (remaining moisture less than 2 wt%). The resulting slurry is pressurized to about 2 MPa and preheated to 400°C before being fed into an extraction reactor (digestor) where most of the organic matter of the coal is extracted by the solvent. No hydrogen or catalyst is used in the extraction stage.

The reactor effluents are cooled and fed into a series of flash separation vessels to separate the lighter gases from the solids-containing slurry stream. Hydrogen is recovered from the gases by cryogenic methods and recycled to the hydrocracker. Light hydrocarbon gases are sent for further processsing in a gas plant.

The slurry is sent to batch filters for removal of solid residues. The wet filter-cake, containing some solvent, is dried under reduced pressure using a 3-stage steam ejector system. The dry, solid residue is mechanically conveyed to the feed hopper of a Shell-Koppers gasifier for production of fuel gas and make-up hydrogen for the process. For Saskatchewan lignites, Koppers-Totzek gasifier would be more suitable.

The filtrate is pressurized to about 20 MPa and mixed with hydrogen before being fed into a fixed-bed, catalytic hydrocracker. The hydrocracker effluents are cooled in a series of heat

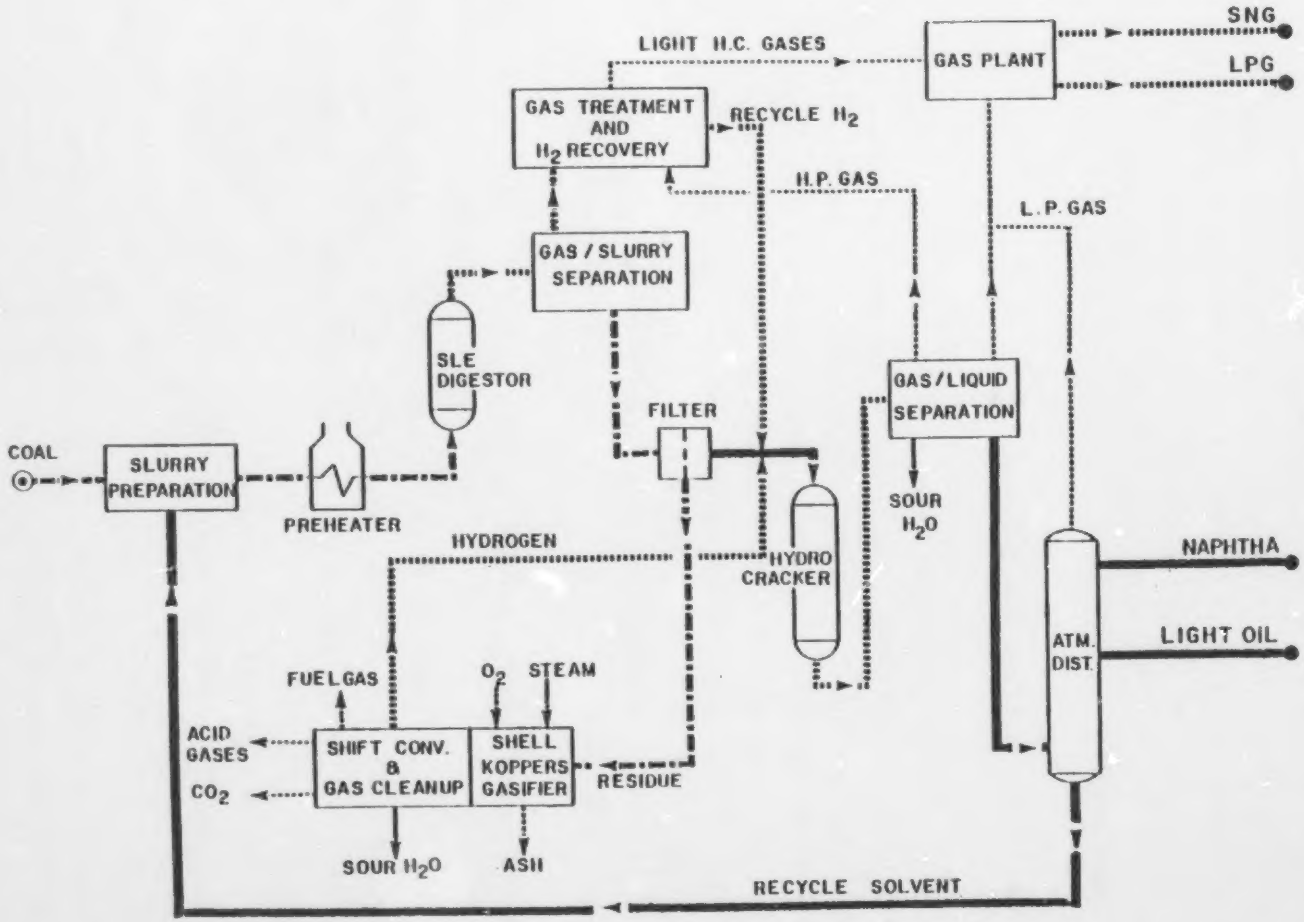


Figure 2-18 Simplified Process Flow Diagram of the LSE Process

exchangers, and separated into high-pressure and low-pressure gas streams, condensed hydrogenated extract and sour-water streams. The condensed hydrogenated extract is preheated and fractionated to yield primary distillate products: naphtha (C_5-200°C), mid-distillate (200-300°C) and heavy oil ($<$ 300°C). The fractionator bottoms ($>$ 300°C) are recycled to the slurry preparation unit.

- **Gas Plant and Light-Ends Recovery**

The gases from various streams are combined, compressed and fed into a deethanizer column. The overhead from the deethanizer is a pipeline gas (SNG). For Estevan and Willow Bunch lignites, all the lighter hydrocarbon gases are used as fuel gas. The deethanizer bottoms are treated in an LPG Merox unit before entering a splitting column where propane and butane are separated. Propane is dried by passing it through a coalescer/salt-drier unit, compressed and sent to storage. Butane is cooled and sent to storage.

- **Hydrogen and Fuel Gas Production**

Hot, dry char (residues) from the coal liquefaction section is crushed in ball mills and then gasified in the presence of steam and oxygen to produce raw synthesis gas which undergoes a subsequent shift-conversion step. Acid gases (H_2S, CO_2) in the shifted gas are selectively removed and the purified hydrogen-rich gas is compressed and sent to the catalytic hydrocracking unit.

A portion of the raw synthesis gas, after acid-gas removal, is used as fuel gas for the process.

2.5.2.3 Product Yield, Distribution and Efficiencies

The product yield and distribution of a self-sufficient, conceptual, commercial LSE plant are presented in Table 2-24 for U.K. bituminous coal (CRC 702) and Saskatchewan lignites (Estevan and Willow Bunch).

The results show that the net liquid-product (C_5-300°C) yield is about 2.7 bbl/t of as-received bituminous U.K. coal compared to 1.3 bbl/t and 1.1 bbl/t of as-received Estevan and Willow Bunch lignites. In other words, the yield is almost 50% and 60% lower for Estevan and Willow Bunch lignites respectively. On MAF basis, the yields are 3.1 bbl/t of U.K. coal, 2.3 bbl/t of Estevan lignite and 2.1 bbl/t of Willow Bunch lignite.

The "overall product" efficiency for bituminous U.K. coal is 71.0% compared to 62% and 61% for Estevan and Willow Bunch lignites, respectively. The "liquid product" efficiency are 57.5%,55.0% and 54% respectively.

2.5.2.4 Suitability/Availability of Process Inputs

- **Reduction of Moisture Content**

The moisture content (35 wt%) of Estevan and Willow Bunch lignites is reduced to 15 wt% by air drying, as suggested by the process developer, before the coal is fed to slurry preparation unit. Feed coal for gasifiers in the utilities section (for steam and power generation) is dried to about 5 wt% moisture.

Residual moisture present in coal to be fed to the extraction reactor is further reduced to less than 2 wt% by a slurry-drying method. Reduction of moisture to these levels for Saskatchewan lignites is not expected to be a problem.

- **Ash Content of Saskatchewan Lignites**

The Estevan and Willow Bunch lignites contain 15.4 wt% and 21 wt% ash, respectively. In the LSE process, solid/liquid separation is carried out using filtration methods. Therefore, problems associated with the pumping and handling of the highly viscous, abrasive, and solids-containing slurry, as in the case of vacuum distillation towers, are avoided. Instead, the residues are mechanically removed from the filter elements and conveyed to the feedhoppers of the gasifier unit using a screw feeder system. Moreover, the residues contain only about 5 wt% solvent and extract. Consequently the loss of coal-derived liquids is also minimal.

- **Availability of Process Inputs**

In the LSE process, the main process inputs are coal, water and catalyst. Requirements of water are about 500 L/s and of hydrocracker catalyst (Co/Mo) about 250,000 kilograms per year. Electricity is generated on site.

2.5.2.5 Product Quality, Utilization and Upgrading Schemes

At NCB, the primary emphasis has been on upgrading of the coal-

Table 2-24 Product Yield and Efficiencies in the LSE Conceptual Plant

(Basis: 25,000 t/d as-received coal)

	U.K. Bituminous (CRC 702)*	Estevan Lignite**	Willow Bunch Lignite**
Coal (as received)			
Moisture, wt%	10.0	35.0	35.0
Ash, wt% MF coal	11.0	15.4	21.0
Coal Fed to Reactor			
Moisture, wt%	5.0	2.0	2.0
Ash, wt% MF coal	11.0	15.4	21.0
Input			
MF Coal to Reactor, t/d	15,050	11,640	11,640
MF Coal to Gasifiers, t/d	-	3,925	3,926
MF Coal to Utilities, t/d	7,470	685	684
Electricity, MW	-	-	-
Ouput, t/d (bbl/d)			
Pipeline Gas (SNG)	1,368	-	-
Propane/Butane (LPG)	876	495	447
Naphtha (C_5-200°C)	2,969 (22,810)	1,770 (12,230)	1,457 (10,069)
Mid-distillate (200-300°C)	5,970 (41,077)	3,049 (19,995)	2,926 (17,161)
Sulfur	114	87	125
Phenol	N/A	N/A	32
Ammonia	245	76	64
Liquid Product Yields (C_5-300°C)			
bbl/t as-received coal to plant	2.7	1.3	1.1
bbl/t MF coal to plant	2.8	2.0	1.7
bbl/t MAF coal to plant	3.1	2.3	2.1
Naphtha in Total Liquid, wt%	33.4	36.7	33.2
Thermal Efficiency, %			
Overall	71.0	62.0	61.2
Liquid Products	57.5	55.0	54.0

Sources:

* NCB Report, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of U.K. National Coal Board", Feb. 1981.

** W.P. Syrett, NCB, Personal Communication to J.A. Mikhlin, SNC Inc., NCB Ref. No. DR 3000, DR 496/2, Oct. 1981; and NCB Rg No. DR 496/2, January 25, 1982.

derived liquids obtained from U.K. bituminous coals to blendstock for gasoline and to diesel fuel.

The coal-derived liquid is an ash-free liquid having a hydrogen content typically 50% higher and a heteroatom content about 25% lower than those of the feed coal. Since no hydrogen is used during extraction, the decrease in oxygen, nitrogen and sulfur content is moderate.

If the extract is directly subjected to distillation, a very small portion (< 7 wt%) can be recovered as a distillate product boiling below 350°C. If heated above 350°C, the extract is unstable. Therefore, catalytic hydrocracking is carried out to reduce the heteroatom content, to increase the hydrogen content, and to produce stable, low-molecular weight distillate products.

Extensive bench-scale experiments to investigate the hydrocracking operation have been carried out at NCB. Kinetic studies using a stirred autoclave system with powdered catalyst show that the conversion of coal-derived extract to light distillates generally follows a first-order chemical reaction. The activation energy is in the range of 125-175 kJ/mole, which is similar to that obtained for hydrocracking of petroleum-derived residues. Thus, it is possible to increase the hydrocracker conversion, not only by the choice of an appropriate catalyst but also by increasing the hydrocracking residence time.

More than fifty catalysts have been tested at NCB in a first selective study, and other catalysts are still being investigated. From these laboratory tests, COMOX 471 (Co/Mo on alumina) gave the best result and has been selected as a hydrocracking catalyst for the proposed pilot-plant operation. The experimental results indicate that between 80% to 90% of the heteroatoms are removed in the distillate products.

The raw naphtha (C_5-200°C) from the distillation unit has a Research Octane Number (RON) of about 80 and is not suitable as high-octane gasoline without further processing. The raw straight-run naphtha is initially hydrotreated mainly to remove sulfur and nitrogen-containing compounds which would otherwise poison the reformer's platinum catalyst. Laboratory-scale experiments on catalytic reforming over a commercial Pt/Al_2O_3 catalyst show that the naphtha may be reformed to a C_6-C_8 aromatic product with RON of about 108. The yield of reformate is about 83 wt% and contains 80 to 85 wt% of the aromatic BTX fraction. The stabilized reformate (RON 108) is blended with butane for volatility adjustment and with raw naphtha (RON 80) to produce a final gasoline product having a RON of 98.

The raw mid-distillate stream is hydrotreated to improve the Cetane number, ignition quality and stability of the product in order to meet diesel fuel specifications. The Cetane number of the diesel product is 45 and meets the ignition requirements for most grades of diesel fuel. Pour point depressants, anti-smoking agents, anti-oxidants and sludge dispersants may be added depending on the grade of diesel fuel required.

2.5.2.6 Areas Requiring Further Development

During the laboratory and bench-scale process development, many problem areas were identified and several modifications have been made in the process for improved and reliable operation. These improvements have also been incorporated in the design of the proposed pilot plant which will have the capacity of processing up to 22.7-t/d of coal. Long-duration pilot plant runs are required not only to establish the design and scale-up data but also to provide a more realistic appraisal of the reliability and operability of the equipment items.

Some of the areas which require further development and demonstration on a large scale are slurry preheating, reactor control and performance, liquid/solid separation, hydrocracking catalyst, product characterization and upgrading, gasification system for solid residues, materials of construction, environmental impact, occupational health and safety, and waste-water treatment.

2.5.2.7 Status and Commercialization Prospect

The National Coal Board has been developing the Liquid Solvent Extraction (LSE) process since 1973. During this period, the process has been tested and demonstrated on a laboratory scale and subsequently in a continuous bench-scale unit having the capacity to process up to 30 kg/h of coal.

The primary emphasis in the bench-scale program has been on the investigation of the influence of operating variables on product yield and quality, optimization of operating conditions, monitoring of equipment and material performance, and in certain process steps, evaluation of alternative modes of operation. Characterization and upgrading of primary distillate products to transportation fuels have also been carried out.

2.5 LIQUID SOLVENT EXTRACTION PROCESS (cont'd)

Different types and ranks of coals have been evaluated, but extensive experimental runs have only been made with high-volatile U.K. bituminous coals. These experiments have been performed under a wide range of operating conditions and sufficient data have been established regarding the influence of coal residence time, coal particle size, temperature, pressure, solvent-to-coal ratio, and type of solvent on extract yield and quality. Residence time, temperature and pressure are the most significant variables, whereas solvent flow rate and coal particle size have a minor influence on the extract yield.

The extract and residual char have been tested to evaluate the potential of these intermediate products in subsequent processing steps, i.e. hydrocracking of coal-derived extract, and gasification and combustion of residual char.

Catalytic hydrocracking of extract has been carried out in a 80-cm^3 "trickle-bed" hydrocracker and then in a continuous 2L fixed-bed hydrocracking unit. More than fifty hydrocracking catalysts have been evaluated, of which COMOX 471 (Co/Mo on alumina) gave the best result. This catalyst will be further tested in the proposed pilot-plant operation.

Gasification tests of residual char have been carried out in 40-cm and 150-cm fluidized-bed units with encouraging results indicating the suitability of these residues as feedstock for the production of make-up hydrogen for the process.

Laboratory-scale tests have been made to upgrade the coal-derived distillates to marketable products. The results indicate that the primary coal-derived products may be upgraded to motor gasoline and diesel fuel. Further work to determine the optimum operating conditions and optimum overall processing scheme is required.

Of the various methods of solid/liquid separation, NCB has selected filtration. Although expensive (10% of the total operating cost), NCB is of the opinion that this method is cost effective and has a potential for further development. Extensive tests (9000 hours of operation) have been made in 30-kg/h continuous filtration units. The detailed design of the filters for the proposed pilot plant has been carried out, but demonstration on a large scale is required. The conceptual design and cost estimation of the filters for a commercial plant is in progress.

To demonstrate the process on a larger scale, the decision was recently made to proceed with a pilot plant (22.7 t/d of coal) for the LSE process. The schedule calls for one-year preconstruction activity, followed by a two-year construction period and a three-year operating and testing period. Process and engineering data for the reliable design and scale-up of a commercial plant will be obtained. NCB expects the process to reach the commercialization readiness stage around 1990.

2.6 LURGI-RUHRGAS PROCESS

The Lurgi-Ruhrgas (L-R) process is a flash carbonization process in which coal is pyrolyzed using product char as a heat carrier. The carbonization is carried out by intimate mixing of char with feed coal in a mechanical screw-mixer which also serves as a pyrolysis reactor. The main products are char, tar and hydrocarbon gases. The quantity of char produced is usually high (50% or more). The tar may be hydrogenated to yield synthetic oil.

The main feature of the process is the circulating of hot carrier, which consists of the fine-grained char produced in the process itself. Another novelty of the process is that char is simultaneously heated and pneumatically conveyed to the carbonizer thus eliminating the need for a separate heater.

2.6.1 History and Development of the Process

The L-R process has been developed by Lurgi Kohle und fur Mineraloltechnik GmbH, in cooperation with Ruhrgas AG company. The process has been used for carbonization of coals, oil shales, tar sands and asphaltic rocks. Cracking of hydrocarbons to produce olefins has also been tested.

The initial development of the L-R process was directed towards production of town gas from coal. The process was initially tested in continuous laboratory units in which the carbonization reactor was a shaft furnace operating with circulating ceramic balls (10 to 12 mm in diameter) as heat carriers. The basic design included a countercurrent flow of descending hot-ceramic balls and coal particles pneumatically transported from the bottom to the top. This system required mechanical conveying and reheating of the ceramic balls causing considerable operating problems.

Later modifications included the use of product char as heat carrier instead of ceramic balls. The modified process was tested in a 12-t/d pilot plant at Herten (Germany) for several years. The process was used for devolatilization of a variety of coals ranging from low-volatile bituminous to lignite. Parallel to the experimental work at Herten, a large pilot plant was operated in close combination with the power house of the Furst Leopold colliery from 1957 to 1960 at Dorsten (Germany) for processing of 240 t/d of high-volatile bituminous coal.

Following the completion of the developmental work on coal carbonization in 1961, two commercial L-R units with a throughput capacity of 1600 t/d of dry lignite were built and commissioned in Yugoslavia in 1963. The plants were built to produce grained char of 15 - 20 wt% volatile-matter content for use as an additive to coking coal in a conventional coke oven works. These plants have been shut down. Two additional L-R units with feed coal capacities of 350 and 700 t/d were built at Bottrop (Germany) and Scunthorpe (Great Britain) for production of char fines from bituminous coals for hot briquetting. These units were operated at carbonizing temperatures around 750 °C.

Renewed interest in the L-R process for partial liquefaction of coal has led Lurgi to build a new test unit with a feed capacity up to 24 t/d at their own premises in Frankfurt/Main (F.R.G). The purpose of the unit is to test hydrocarbon feedstocks of various types and origins in addition to coal such as tar sands, asphaltic rocks, oil shales, etc.

2.6.1.1 Overall Conceptual Process Configuration

A simplified conceptual process configuration of the L-R process is shown in Figure 2-19. The diagram consists of four major process sections which are described below:

- **Coal Preparation**

In the coal preparation unit, run-of-mine coal is reduced in size to about 5 mm, washed to control the ash content and sent to the feed bins of the carbonization section.

- **Coal Carbonization**

The coal carbonization section consists of a lift pipe, char separation/collection bin, a mixer/carbonizer (pyrolysis reactor), surge bin and a cyclone separator. The feed coal enters the mixer where it is intimately mixed with hot char at about 700°C from the collecting bin. The coal-to-char mixing ratio is kept between 4 and 8. This is chosen to maintain the mixing temperature of approximately 600°C.

Most of the pyrolysis takes place in the mixer/carbonizer where the residence time of the coal is very short. Due to the short residence time, secondary decomposition reactions are suppressed, thus allowing a high liquid product yield. The gases and condensible vapors leaving the mixer/carbonizer are sent to a cyclone where the entrained coal and char particles are removed and returned to the circulation system downstream of the surge bin. The gases/vapors are sent to the gas/liquid separation section.

The freshly produced char and the cooled heat-carrier (recycled char) are discharged into a surge bin from where the solids are withdrawn into the lower part of the lift pipe and pneumaticlly conveyed by hot combustion gases. The hot gases are produced in the lift pipe itself by partial combustion of char with preheated air. The hot combustion gases at above 700°C are fed into the char separation and collection bin. Fine char particles are removed in the cyclone and returned to the carbonizer/mixer. The flue gases are sent to a waste-heat recovery system (not shown in the diagram).

Particulate char collected in the lower part of the separation/ collection bin flows continuously as a compact bulk material into the mixer/carbonizer to complete the cycle. The circulation of char as the heat carrier is controlled by gate valves in the conduits between the collecting bin and the mixer, and the surge bin and the lift pipe.

Surplus char is continuously withdrawn from the bottom of the surge bin and sent to the storage area.

- **Gas/Liquid Separation**

Gas/liquid separation is carried out by condensing the vapors in stages in order to obtain separate fraction of liquid products and also to avoid formation of undesirable oil/water emulsions.

The first stage condensation is done in a uniflow scrubber where the effluents are cooled and at the same time, the fine coal/char particles not separated in the cyclone are removed by circulating heavy tar. The fine solid particles collected in the tar fraction are removed by centrifugation.

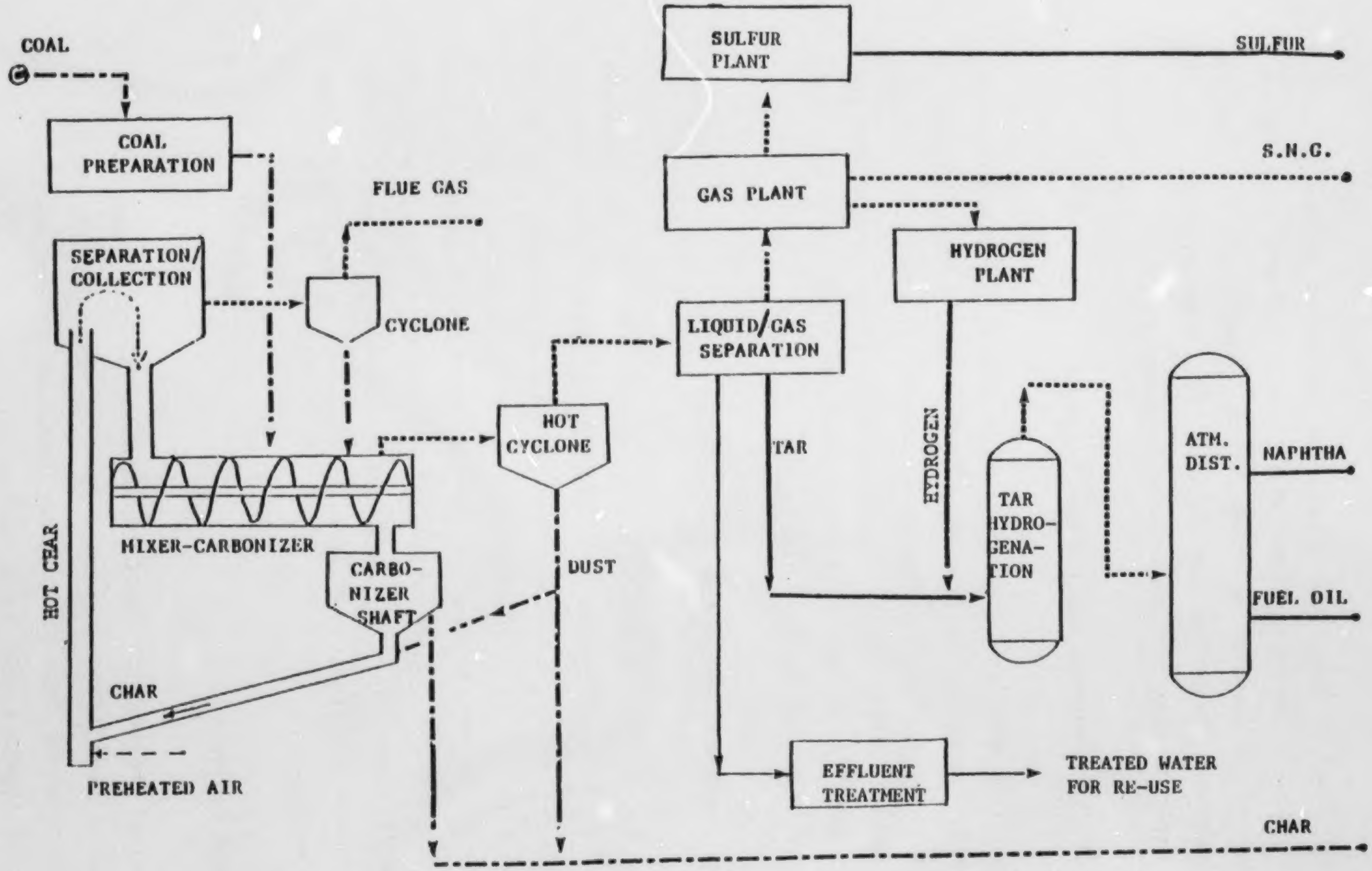


Figure 2-19 Simplified Flow Diagram of Lurgi-Ruhrgas Process

Source: P. Nowacki, "Coal Liquefaction Processes", Chemical Technology Review No. 131, Energy Technology No.45, Noyes Data Corp., N.J. 1979

In the final condensation stage, the gas/vapor stream is cooled to 30-40°C. Tar and water which are condensed together and passed to a separating tank where separation is done by decantation. Recovered tar is sent for hydrogenation and the sour water is sent for waste-water treatment.

- **Tar Hydrogenation and Product Recovery**

The tar from gas/liquid separation is highly viscous and is not an easily transportable product. It is solid at room temperature and would require hydrogenation to render it a synthetic crude. The synthetic crude may be fractionated to yield naphtha and fuel oil. Details of the tar hydrogenation step have not been reported by the process developer.

- **Gas Treatment and Hydrogen Generation**

In the gas plant, carbonization gases are treated for acid-gas removal. The acid gases are sent to a sulfur plant for the recovery of elemental sulfur. The purified hydrocarbon gas has a heating value of 26 - 32 MJ/m^3 and may be methanated to pipeline quality SNG or may be used to produce hydrogen required for the tar hydrogenation. Alternatively, hydrogen may also be produced by gasification of char (not shown in the diagram). For Saskatchewan lignites, hydrogen is assumed to be produced by the latter method.

2.6.2 Technical Review

A technical review of the Lurgi-Ruhrgas process based on the information available in the open literature and material directly received from the process developer in conjunction with processing lignitic coal is presented below:

2.6 LURGI-RUHRGAS PROCESS (cont'd)

2.6.2.1 Types of Coal Investigated

The process has been tested for various types of coal ranging from lignite to low-volatile bituminous coals. Typical analyses of some of these coals are presented in Table 2-25. The subbituminous coals of German and Australian origin have been processed on pilot plant le[illegible] Typical analysis of a German subbituminous coal which has been extensively tested in the pilot plant is given in Appendix B. A commercial plant has been operated with lignites in Yugoslavia.

2.6.2.2 Product Yields, Distribution and Efficiencies

In the L-R process, the direct contact between hot char and coal results in an extremely fast heat-transfer rate in the order of 4000 $kJ/m^2h°C$. Due to the flash heating there are considerable shifts in the yield and quality of the carbonization products. However, this effect can be greatly obscured by secondary reactions, such as tar-vapor cracking. In practice, this means selecting the lowest possible carbonization temperature which must, however, be high enough to ensure maximum tar yield. Furthermore, the carbonization gases and condensible vapors must be quickly withdrawn from the hot zone of the retort and quenched immediately to stabilize the products.

Figure 2-20 shows the dependence of product yields on temperature. The data shown are obtained from numerous tests on various types of coals of volatile matter 19 to 45 wt% MAF basis. The results indicate that a carbonization temperature of around 600°C is optimum in terms of tar yields. For comparison purposes, Fischer Assay data at 600°C is also shown by a vertical column (designated as "F") in Figure 2-20. The actual tar yields from the L-R process have been found to be 129-212 wt% of the tar yields obtained by Fischer Assay.

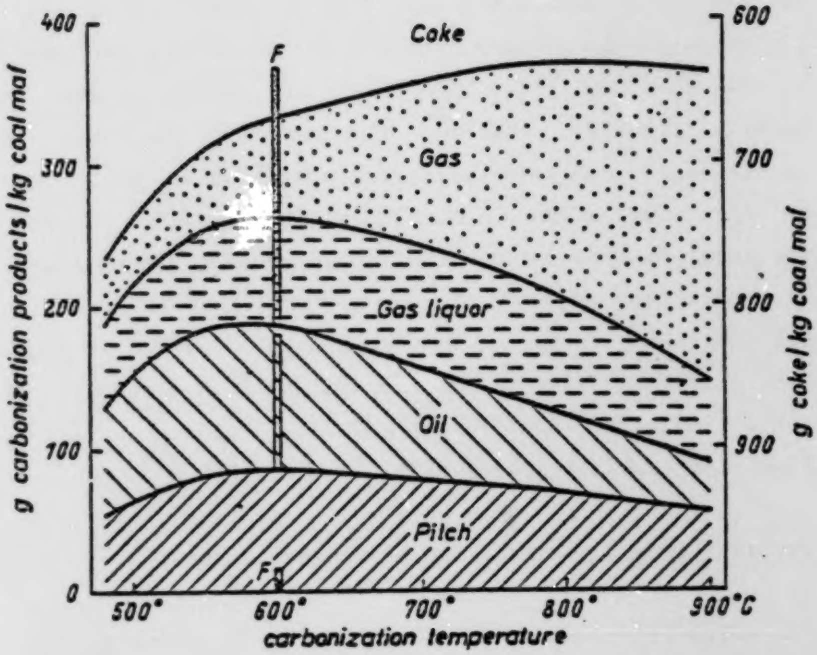
Table 2-25 Typical Analyses of Some of the Coals Tested in the L-R Process

	Sibbituminous			Lignite		
	German	Waloon Australia	West Moreton Australia	Bosnia Yugoslavia	German	Qeenland Australia
Moisture, wt%	8.9	9.1	1.7	36.0*	N/A	13
Ash, wt% MF Coal	4.9	27.6	23.8	8.6	N/A	17
Tar Content, wt% MAF coal (by Fischer Assay)	11.3	35.2	19.2	11.9	21.9	N/A

* as-received basis

Sources: R.W. Rammler, "The Lurgi-Ruhrgas Process for Carbonization of Coal" paper presented at the EPRI Conference on Coal Pyrolysis, Palo Alto, CA, Feb. 25-26, 1981.

D.N. Perkins, "Production of Liquid Hydrocarbons from Coal and its Impact on the Electricity Supply, Industry in Queensland", Proceedings of the EPRI Conference on Coal Pyrolysis, Palo Alto, CA, Feb. 25-26, 1981.



Note Vertical Column designated by "F" represents data obtained by Fischer-Assay

Figure 2-20: Lurgi-Ruhrgas Carbonization Product Yields

Source: R.W. Rammler, "Retorting of Coal, Oil Shale and Tar Sand by Means of the Lurgi Ruhrgas Process", German-Canadian Coal Refining Symposium, Edmonton, Alberta, April 20-21, 1978

Typical product yields from a subbituminous coal (11.3 wt% tar by Fischer Assay) are presented in Table 2-26. The liquid yield is 21 wt% MAF coal, which corresponds to 188% of the Fischer Assay tar yield.

Estimated product yields for a conceptual Lurgi-Ruhrgas plant operating with Estevan lignite are presented in Table 2-27. Fischer Assay of 10 wt% and liquid product yield of 190% of Fisher Assay was assumed. German subbituminous coal was taken as a reference coal (see Table 2-26) to predict product yield for Estevan lignite. Make-up hydrogen and additional utilities requirement for a self-sufficient operation is met by utilization of char. The amount of char shown in Table 2-27 is about 70% of the total char produced. The liquid product yields and efficiencies have been adjusted for zero electricity import using a fuel oil equivalent factor of 6MM Btu/bbl (6.33 GJ/bbl). The adjusted liquid product yields are 1.0 bbl/t MF coal or 1.2 bbl/t MAF coal. Similarly, the overall and liquid product efficiencies are 72.4% and 31.4%, respectively.

2.6.2.3 Suitability and Availability of Process Inputs

The suitability of the Lurgi-Ruhrgas process for different types of caking and non-caking coals has been demonstrated at various levels of operation from 240-t/d pilot plant to 1600-t/d commercial plant. The ash and moisture content in subbituminous and lignitic coals varied from 5 to28 wt% and 2 to 36 wt% respectively. Nevertheless, the process suitability must also be viewed on the basis of desired end-products. If char is considered as a main product, then the process is suitable for most types of coal. On the other hand, if the liquid products are desired as one of the main products, the preliminary judgement on

Table 2-26 Typical Product Yield in the L-R Process for a German Subbituminous Coal

	German Subbituminous
Moisture wt% of feed coal	8.9
Volatile Matter, wt% MF coal	38.5
Ash, wt% MF Coal	4.9
Tar Yield (Fischer Assay), wt% MAF coal	11.3
Operating Conditions	
Coal Feed Rate, t/h	0.5
Carbonization Temperature, °C	600
Product Yield, wt% MAF coal	
Gases	13.9
Tar*	21.3
Water	15.1
Char	49.7
	100.0

* 188% of Fischer Assay

Source: R.W.Rammler, "Practical Experience with the Lurgi-Ruhrgas process for the Devolatilization of Fine-Granular Fuel", translation of an article published in Erdol und Kohle, Erdgas, Petrochemie, 19, pp. 117-121, 1966.

process suitability may be made on the basis of Fischer Assay tar yields of the feed coal. For instance, coals with Fischer Assay tar yield less than 10%, as indicated by the process developer, should not be considered as suitable.

The Fischer Assay of both Estevan and Willow Bunch lignites, as determined by B.C. Research Laboratories, are shown in Table 2-28. The Estevan lignite has almost 10 wt% Fischer Assay tar yield but the Willow Bunch lignite showed only 6.0 wt% tar yield. Char yield in both cases is in the range of 65-70 wt% MAF coal.

In Table 2-27 , the estimated amount of net liquid products is calculated as 190 kg/t MAF coal. Net char yield is 365 kg/t MAF coal but the total char yield is actually about 525 kg/t MAF coal. Thus, it can be seen that the overall suitability of the process is related to the marketability of large quantities of char. Char combustion by large power plants may be viewed as one of the ways to utilize the char product.

In terms of process inputs, the L-R commercial plant operating with Estevan lignite will require about 54 MW electricity and 190 L/s as make-up water.

2.6.2.4 Product Quality and Upgrading Required

Table 2-29 presents the API gravity, TBP distillation and the elemental analysis of naphtha, mid-distillate and heavy oil produced from a German subbituminous coal and a Yugoslavian lignite. The elemental analysis reveals that the liquid products from both coals contain high levels of nitrogen, sulfur and oxygen. Severe hydrotreating (over 70 m^3 H_2/bbl) will be required to reduce the heteroatoms content to the desired level.

Table 2-27 Estimated Product Yield and Efficiencies of a L-R Conceptual Plant
(Basis: 25,000 t/d as-received coal)

	Estevan Lignite
Coal (as received)	
Moisture, wt%	35.0
Ash, wt% of MF coal	15.4
Coal Fed to Reactor	
Moisture, wt%	2.0
Ash, wt% of MF coal	15.4
Input	
MF Coal to Reactor, t/d	16250
MF Coal to Gasifier, t/d	-
MF Coal to Offsites, t/d	-
Electricity, MW	54
Output, t/d (bbl/t)	
Fuel Gas	1710
Light Oil (C_5-200°C)	1526 (11187)
Fuel Oil (200-525°C)	1107 (7579)
Char*	5050
Sulfur	34
Liquid Product Yield**	
bbl/t as-received coal to plant	0.66
bbl/t MF coal to plant	1.02
bbl/t MAF coal to plant	1.20
Naphtha in Total Liquid, wt%	58.0
Thermal Efficiency, %	
Overall	72.4
Liquid Products	31.4

* About 30% of the total char produced is used for make-up hydrogen and additional utilities production. The value shown here is the net char yield.

* Liquid product yields, expressed as bbl/t of coal to the plant, has been adjusted to zero electricity import using 10,000 Btu/kWh (10,500 kJ/kWh) and a fuel oil equivalent factor of 6 MM Btu/bbl (6.33 GJ/bbl).

Source: R.W. Rammler, "The Lurgi-Ruhrgas Process for Carbonization of Coal" paper presented at the EPRI Conference on Coal Pyrolysis, Palo Alto, CA, Feb. 25-26, 1981.

Table 2-28 Fischer Assay of Saskatchewan Lignites as Determined by B.C. Research Laboratories

(Carbonization Temperature = 500°C)

	Estevan Lignite		Willow Bunch Lignite
	Untreated	Predried to 10 wt% H_2O	Untreated
Fischer Assay Yield, wt% MAF coal			
Char	65.0	69.4	65.8
Tar	9.4	9.4	5.8
Water Formed	10.1	9.6	15.5
Total Gas (by difference)	15.5	11.6	12.9
Composition of Gas, vol%			
CH_4	17.5		13.8
C_2H_6	3.6		2.2
H_2	28.0		44.9
CO	9.0		4.2
CO_2	41.9		34.9
H_2S	0.1		0.1

Source: B.C. Research Laboratories,"Liquefaction and Pyrolysis Tests of Samples of Estevan and Coronach Lignites", Project Report 2-02-691, 1982.

Table 2-29 Properties of Liquid Products from the L-R Process

	Subbituminous Coal			Lignite		
	Naphtha	Mid-distillate	Heavy Oil	Naphtha	Mid-distillate	Heavy Oil
Sp. Gravity, °API	60.2	17.5	-19.1	50	27.1	2.8
Distillation, °C						
IBP	33	85	-	45	85	225
5%	47	164	-	-	-	-
50%	74	253	-	-	-	-
95%	138	315	-	-	-	-
FBP	139	318	-	182	296	471
Elemental Analysis wt%						
Carbon	N/A	83.3	83.1	84.6	84.6	83.0
Hydrogen		10.0	6.1	10.3	10.3	9.0
Nitrogen		0.3	1.0	0.6	0.5	0.8
Sulfur		0.4	0.4	0.8	0.9	0.5
Oxygen		6.0	9.4	3.7	3.7	6.7

Notes: Products from subbituminous coals obtained from 240 -t/d pilot plant at Dorsten, W. Germany
Products from Lignite obtained from 1600 -t/d commercial plant, Lucavae, Yugoslavia.

Sources: R.W. Rammler, "Practical Experience with the Lurgi-Ruhrgas process for the Devolatilization of Fine-Granular Fuel" Translation of an article published in Erdol und Kohle, Erdgas, Petrochemie, 19, 117-121 (1966).

R.W. Rammler, "The Lurgi-Ruhrgas Process for Carbonization of Coal" Paper presented at the EPRI Conference on Coal Pyrolysis, Palo Alto, CA, Feb. 25-26 1981.

The process developer has not yet carried out studies pertaining to liquid-product upgrading and their end-use compatibility. Therefore, at this stage, no assessment can be made regarding upgrading of liquid products and marketability of the end-products.

Typical ultimate and proximate analysis and grain size distribution of the total char (including dust from cyclones) produced by carbonization of German subbituminous coal at 590°C is given in Table 2-30. The volatile matter content of the char is about 6.4 wt% on MF basis. In general, char of about 7.0 wt% volatile matter has been produced with various coals at the carbonization temperature of 600°C. Ash content of the char shown in Table 2-30 is twice that of the feed coal. Similar, doubling of ash has been observed for other coals tested in L-R process.

Another important characteristic of the char is its sulfur content which has an impact on direct utilization of char and the subsequent processing steps. Typically, sulfur content of the char obtained in L-R process ranges from 50% to 65% of the feed coal. In the case of Estevan lignite (0.6 wt% sulfur on MF basis) the char produced would have a low sulfur content, and as such, the char could be utilized with minimal gas treatment for SO_2 removal.

The heating value of the char in the L-R process is typically equal to or somewhat higher than the feed coal in case of the higher-rank coals. For lower-rank coals or those with higher ash content, the heating value is generally lower (up to 85% the feed coal). However, it should be noted that in view of the flexibility of the L-R process, the heating value of char may be varied

Table 2-30 Typical Analysis of Char obtained from German Subbituminous Coal

Proximate Analysis, wt% MF basis	
Volatile Matter	6.4
Fixed Carbon	82.8
Ash	10.8
Ultimate Analysis, wt% MF bais	
Carbon	82.7
Hydrogen	1.6
Nitrogen	1.1
Sulfur	0.4
Oxygen	3.4
Ash	10.8
Grain Size Distribution, wt%	
< 0.10 mm	12.4
0.10-0.20	12.2
0.20-0.315	10.6
0.315-0.40	9.1
0.40-0.50	11.5
0.50-0.63	8.3
0.63-0.80	10.7
0.80-1.00	11.0
1.00-1.25	4.2
1.25-1.60	4.7
1.60-2.00	3.5
> 2.0	1.8

Source: R.W. Rammler "The Lurgi-Ruhrgas Process for Carbonization of Coal", paper presented on Coal Pyrolysis, Palo Alto, CA, Feb. 25-26, 1981 through optimization of the operating conditions in the reactor; to have a heating value equal to or higher than the feed coal.

The size distribution of the char shown in Table 2-30 appears to be somewhat too coarse for conventional pulverized firing in the power boilers. Further pulverization would be required. In case of low-rank coals, the char has a tendency for particle disintegration. Furthermore, pneumatic conveying in the lift pipe provides to a certain extent grinding effect. This mainly occurs in the bottom section of the lift pipe where the particles are accelerated. By changing the conveying speed and the flow rate of the circulating hot char, the size distribution of the final char product may be influenced within the operational limits.

From above, it is evident that the quality of both liquid products and char is dependent on many factors. This only serves to stress the importance of determining the characteristics of these products through actual experimental tests. For a particular feed coal, the impact on the equipment design and its operability may be substantial.

2.6.2.5 Status and Commercialization Prospects

The L-R process is undoubtedly the most proven from the available coal pyrolysis technologies. Four plants, the largest having a feed rate of 1,600 t/d of dry lignite have been built by Lurgi. On this basis, it would perhaps appear, that the Lurgi-Ruhrgas process is commercially proven but not up to the size required to process 25,000-t/d "as-received" (16,250 t/d dry coal basis) of Saskatchewan lignites. The 1,600 -t/d plant may be scaled up to a feed rate of about 3,000 t/d of coal. Thus, the full-size plant can be built having six parallel-trains.

The carbonization section of the L-R plant can be regarded as a proven commercial operation. Similarly, the hydrogen production plant using pyrolysis gas is also commecially proven for many years in the chemical industry notably in production of synthetic ammonia.

Tar hydrogenation, however, has not been carried out commercially. The major problems foreseen are associated with the unstable nature of the tar and the risk of polymerization with subsequent coke formation in the equipment items. This would require additional development work which has yet to be carried out. At the same time, it must be noted, that the process could be considered for economic commercial application only under the condition that the char would be fully utilized.

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3.0 TECHNO-ECONOMIC ASSESSMENT OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS

3.0 TECHNO-ECONOMIC ASSESSMENT OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS

A preliminary evaluation of five "direct" coal liquefaction processes and one pyrolysis process, presented in the previous chapter, was carried out for designated Estevan and Willow Bunch lignites from Saskatchewan, Canada. The details and results of the process evaluations are presented in Volume I, "Summary and Process Comparisons". Among the "direct" coal liquefaction processes, the LSE process was found to be the most suitable one for Saskatchewan lignites according to the "Base Criteria" set up by the Study Committee.

The LSE process has a high liquid yield and thermal efficiency, meets the product criteria, and offer prospects for commercial readiness around the targetted time period of 1990. High liquid yields are possible from the LSE process for high ash content Saskatchewan lignites because only a relatively small amount of liquid product is lost in the residual char stream as the solid/liquid separation is done by a filtration method. The other "direct" coal liquefaction processes employ hydroclones and/or vacuum distillation methods, consequently a large amount of recoverable liquid product is lost in the residual slurry. The reduction in net liquid product yield results in a corresponding decrease of the thermal efficiency, particularly for Willow Bunch lignite which has a higher ash content (21.0 wt% MF coal) than Estevan lignite (15.4 wt% MF coal).

For further evaluation of the LSE process, a techno-economic assessment has been carried out for a self-sufficient, integrated, commercial plant having a capacity of processing 25,000 tonnes per day of Willow Bunch lignite. Typical proximate and ultimate analyses (average of 206 samples), ash distribution and elemental analysis (average of 133 samples) are given in Appendix C. Proximate and ultimate analyses of the "best" and "worst" samples of Willow Bunch lignite are also included.

The assessment work presented in this chapter has been prepared based on information and data received from the Coal Research Establishment, National Coal Board, England. A confidential information package entitled "The Processing of Saskatchewan Lignite by the NCB's Liquid Solvent Extraction Process: Willow Bunch Lignite", (NCB Ref. No. DR496/2, 7/3, January 23, 1982) is attached in confidential Volume V, "Process Developers' Data and Information".

In addition, a sample of Willow Bunch lignite (Coronach #3216) was sent to NCB for preliminary characterization in the microautoclave unit of the LSE process. Proximate, ultimate and petrographic analyses of the coal sample are given in Appendix D. The results and details of the experimental tests, received as confidential report, "Assessment of a Canadian Lignite for the National Coal Board Liquid Solvent Extraction Process", (NCB Ref. No. DR496/3, April 1982), is attached in confidential Volume V, "Process Developers' Data and Information". However, a summary of the experimental results is presented in Appendix D.

On the basis of the "average" coal analysis given in Appendix C, NCB by comparison with other similar coals extensively tested in the LSE process, predicted that a extraction efficiency up to 82% on MAF basis may be possible. However, in the microautoclave tests, an extraction efficiency of 65% on MAF basis was obtained. The low extraction efficiency may be attributed to the high inertinite content (22 wt%) in the particular sample as compared to the average value (18 wt%). Consequently, the values 65% and 82% may be considered as the "lower" and "upper" levels of the extraction efficiency attainable for Willow Bunch lignite in the LSE process. Therefore, the techno-economic assessment has been presented for these two cases.

The technical assessment of the LSE process includes:

- Overall configuration and conceptual design of a self-sufficient plant processing 25,000 t/d of as-received Willow Bunch lignite.
- Overall material balance, product efficiencies, and utility summary.
- Detailed description of the coal liquefaction section, the characteristic part of the LSE process.
- Product analysis and upgrading schemes of primary liquid products to transportation fuels (gasoline, jet fuel, and diesel).
- Environmental consideration.

The economic assessment of the LSE process includes:

- Capital cost estimation and total investment requirement for a plant located in Saskatchewan.

3.0 TECHNICO-ECONOMIC ASSESSMENT OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS (cont'd)

- Annual operating costs (fixed and variable).

- Product cost estimation based on "Energy Cost" basis and "Product Value" basis.

3.1 OVERALL CONFIGURATION OF A CONCEPTUAL, SELF-SUFFICIENT LSE PLANT

A list of processing areas or sections of a conceptual, self-sufficient LSE plant is presented in Table 3-1. The interrelationship between the various processing areas for both cases (65% and 82% extraction efficiencies) is given in Drawings 3-1 and 3-2 respectively.

An outline of these processing areas was presented in the previous chapter (see Section 2.5.2.2).

A more detailed description along with a summary of thermal efficiency calculations and utilities balance for processing of Willow Bunch lignite is given in the following pages.

Table 3-1 Processing Areas of the Liquid Solvent Extraction Commercial Plant

Section No.	Description
100	COAL PREPARATION
	Storage and Reclaiming
	Crushing and Drying
200	COAL LIQUEFACTION
	Coal/Oil Slurry Preparation
	Slurry Drying and Preheating
	Coal Extraction
	Gas-Vapor and Slurry Separation
	Gas Treatment and H_2 Recovery
	Solid/Liquid Separation (Residue Separation)
	Hydrocracking (Hydrogenation of Extract)
	Gas/Liquid Separation
	Fractionation
300	GAS TREATMENT AND LIGHT-ENDS SEPARATION
	Compression Facilities
	Acid Gas Removal (Amine Plant)
	De-ethanizer
	LPG Splitter
	Gas Drying Facilities
400	HYDROGEN AND FUEL GAS PRODUCTION
	Gasification
	Shift Conversion
	Rectisol Acid Gas Removal and Methanation
	Hydrogen Compression
500	OXYGEN PLANT
600	EMISSION CONTROL
	Claus Sulfur Recovery Plant
	SCOT Tail-Gas Processing Unit
700	EFFLUENT CONTROL
	Sour Water Treatment
	Phenol Recovery
	Ammonia Recovery

Table 3-1 Processing Areas of the Liquid Solvent Extraction Commercial Plant (cont'd)

Section No.	Description
800	PRODUCT STORAGE AND SHIPPING *
900	UTILITIES Raw Water Treatment and Distribution Cooling Tower and Cooling Water System Boiler Feedwater Preparation Condensate Treatment Steam Generation and Distribution Electricity Generation and Distribution Fuel Gas Purification and Distribution Plant Air, Instrument Air and Inert Gas System Flare System
1000	OFFSITES *
1100	PRODUCT UPGRADING AND REFINING Hydrotreatment Reforming Product Blending

* Not shown in the block diagrams (Drawings 3-1 and 3-2)

3.1 OVERALL CONFIGURATION OF A CONCEPTUAL, SELF-SUFFICIENT LSE PLANT (cont'd)

3.1.1 Coal Preparation (Section 100)

The coal preparation section includes coal handling and storage facilities along with coal grinding, classifying and drying equipment to provide coal of desired specification for the following processing areas:

- Coal extraction reactors (Section 200): 70% minus 0.075 mm size, and 15 wt% moisture content
- Koppers-Totzek gasifiers for fuel gas and make-up hydrogen production (Section 400): 70-90% minus 0.075 mm size, and 8-10 wt% moisture content
- Coal-fired boilers for steam and power generation (Section 900): no size reduction but drying up to equilibrium moisture content of 28.5 wt% using low-grade energy (e.g. air drying) that would otherwise be rejected.

The as-received Willow Bunch lignite of approximately minus 2-cm size and containing about 35% moisture is supplied to the plant battery limits by rail cars or by conveyors. In the coal receiving station, the rail cars dump the coal into hoppers located below the track. Coal is withdrawn from the hoppers by means of vibrating feeders and discharged onto a system of belt conveyors. The conveyor system includes facilities for coal sampling, weight-recording and removal of tramp iron by magnetic separators.

Under normal conditions, the coal is conveyed to two coal storage silos, one located within the coal preparation section and the other near the offsite facilities. Surplus incoming coal is conveyed to an adjacent dead-storage area where two stock piles, each of approximately 400,000 tonnes capacity, are prepared to meet 30-day coal requirements. During emergency or shortfall operation, coal may be reclaimed by a travelling-wheel stacker/reclaimer of 1200-tonnes per hour capacity and sent to the coal silos by the reversible conveyer system. In the possible event of a problem with a reclaimer, coal may be retrieved by means of mobile equipment through an emergency reclaiming hopper.

The raw coal from the storage silo in the coal preparation section is transferred to feed-hoppers supplying coal to the bowl mills where simultaneous crushing and drying is carried out. A pneumatic conveyor system utilizing hot gas at approximately 205°C is used to dry and convey the pulverized coal to cyclone classifiers located above the mills. The hot gas is generated by burning process-derived fuel gas in a direct combustion heater. The dried coal from the cyclones is collected in coal storage bins.

3.1.2 Coal Liquefaction (Section 200)

Dried and pulverized coal (70% minus 0.075 mm, 15 wt% moisture) from Section 100 is continuously received in the slurry preparation unit where it is mixed with recycle oil (300 - 370°C) in an oil-to-coal ratio of 2.0. During mixing, the moisture content of the coal is reduced from 15 wt% to less than 2 wt% by a slurry-drying method.

The resulting slurry is pressurized to 2 MPa, preheated to 420°C, and fed into extraction reactors. A residence time of 60 minutes is provided to achieve the required extraction. No hydrogen or catalyst is used in the extraction stage.

The reactor effluents are separated into gas/vapor and slurry streams. The gas/vapor phase is subjected to a sequence of cooling, condensing and separation stages to yield liquid hydrocarbons (condensate), hydrocarbon gases, and sour-water streams. The condensate is sent for fractionation. The hydrocarbon gas stream is treated for removal of acid gases before being sent for LPG processing in Section 300. Sour water is sent to the effluent control unit in Section 700.

The slurry is subjected to batch-filtration to remove the unreacted coal and ash (termed as residual char or residues) from the coal-derived extract. The residues are sent to Section 400 for make-up hydrogen production.

The extract (filtrate) is fed into a catalytic hydrocracker operating at 420°C and 20 MPa. The hydrocracker effluents are separated into high-pressure and low-pressure gas stream, liquid hydrocarbons and sour water streams by cooling and flash separation methods. The liquid hydrocarbon stream is fractionated to yield naphtha (C_5-200°C), mid-distillate (200 - 300°C), and heavy oil ($>$300°C). A portion of heavy oil is recycled to the slurry preparation unit.

The high-pressure sour-gas stream containing unreacted hydrogen is initially treated in an amine unit to remove the acid gases and then cryogenically purified to obtain a hydrogen-rich stream (-96 vol% H_2) which is recycled to the hydrocracker. The spent di-ethanolamine (DEA) is regenerated to yield acid gases which are sent for sulfur recovery in Section 600. The hydrocarbon gases (C_1-C_4) and low-pressure gases are further processed for gas treatment and light-ends recovery.

Further details of the coal liquefaction section with process flow diagram are given in Chapter 3.3.

3.1.3 Gas Treatment and Light-Ends Separation (Section 300)

The low-pressure sour-gas stream from Section 200 is compressed to 2 MPa and combined with gases from Section 1100 to be processed in an amine plant for removal of acid gases using di-ethanolamine as a treating agent. The acid gases are sent to a Claus plant (Section 600) for sulfur recovery, whereas the treated gases are combined with the sweet-gas stream and fed into a de-ethanizer column.

The overhead product ($< C_2$) from the de-ethanizer is used as fuel gas for the plant. The de-ethanizer bottoms are treated in an LPG Merox unit before being fed into an LPG splitter to separate propane and butane. Propane is dried by passing it through a coalescer/salt-drier before being sent to storage (Section 800). The stream of butane is blended with the gasoline product and the remainder is sent to storage (Section 800).

3.1.4 Hydrogen and Fuel Gas Production (Section 400)

The hot char (residues) from the coal liquefaction section and pulverized coal from the coal preparation unit are gasified to produce synthesis gas for the production of fuel gas and make-up hydrogen.

The hot, dry residues are initially pulverized to the desired size (70-90% minus 0.075 mm) in ball mills. The pulverized char

3.1 OVERALL CONFIGURATION OF A CONCEPTUAL, SELF-SUFFICIENT LSE PLANT (cont'd)

and coal are then fed into atmospheric, entrained-type, Kopper-Totzek gasifiers where they react with a "blast" of oxygen and steam in a flame-like reaction. The flame temperature reaches up to 2000°C but the reactor outlet temperature drops to about 1500°C due to endothermic reaction and heat losses. The pressure inside the gasifier is maintained at around 105 kPa. A carbon conversion of up to 99% is obtained.

Most of the ash is withdrawn from the reactor as "molten slag" and quenched in a water-filled compartment to solid ash for disposal. The gas leaving the gasifiers is cooled by water spray to solidify fly ash. Further removal of entrained particles is carried out in electrostatic precipitators where the dust content of the raw gas is reduced to 0.2 mg/m^3. The solid-free gas then passes through waste-heat boilers for generation of high-pressure steam. Additional high-pressure steam is also generated by recovering process heat in the reactor jacket of the gasifier.

The synthesis gas is then treated in a Rectisol unit where sulfur-containing gases (COS, H_2S) are selectively removed using CO_2-rich, sulfur-free methanol as solvent. The sulfur impurities are reduced to less than 1 ppm. A portion of the treated gas (about 15% of the total) is sent to the utilities section from where it is distributed as fuel gas for the process.

The remaining synthesis gas undergoes a two-stage shift-conversion step where CO and steam are reacted to form CO_2 and H_2. The shifted gas, containing mainly H_2 and CO_2, is treated in the

Rectisol unit where it is initially scrubbed with CO_2-rich methanol solvent to remove traces of sulfur impurities and then with lean solvent for removal of CO_2.

The purified hydrogen-rich gas stream (97 vol% H_2) is further treated in a fixed-bed, Ni-catalyst reactor where CO is converted to methane. The hydrogen-rich gas is then sent to the coal liquefaction section.

3.1.5 **Oxygen Plant (Section 500)**

High-purity oxygen (98.0 mol%) required for char gasification is produced in a conventional cryogenic oxygen plant. Oxygen obtained from the plant is compressed and fed into the Kopper-Totzek gasifiers in Section 400.

High-purity nitrogen, which is also produced, is used for the transportation of solids, inert blanketing in liquid products storage vessels and for instrumentation purposes. Low-purity nitrogen is vented to the atmosphere.

3.1.6 **Emission Control (Section 600)**

The acid gas streams from the Rectisol unit (Section 400), amine plant (Section 300) and sour-water stripper (Section 700) are treated in the Claus sulfur plant for recovery of elemental sulfur.

About one-third of the total acid gas stream entering the Claus plant is oxidized with air in a furnace to convert the H_2S to

SO_2. This SO_2-containing gas stream is reacted with the remaining two-third acid gas stream in a two-stage Claus converter to form elemental sulfur ($SO_2 + 2H_2S = 2H_2O + 3S$). The recovered molten sulfur is sent to a storage pit (Section 800) and the off-gases are further treated in a Shell-Claus-Offgas-Treating (SCOT) unit to reduce the sulfur content to an environmentally acceptable level.

In the SCOT process, all sulfur compounds present in the tail-gas stream are selectively converted to H_2S over a cobalt-molydenun catalyst at 300°C in the presence of a reducing gas. The reducing gas is normally CO and/or H_2 and a sufficient quantity of it is present in the tail-gas stream. The exit gases from the SCOT unit, containing 2 to 3% H_2S, are treated in an amine absorber for removal of the H_2S impurity. Treated gases from the absorber are sent to the incenerator stack. The SO_2 content in the vent gases is less than 100 ppm.

The spent amine is stripped to produce an H_2S-rich gas stream which is recycled to the Claus plant for sulfur recovery. The regenerated amine is recycled to the absorber.

3.1.7 Effluent Control (Section 700)

Sour water streams from Sections 200, 400, and 1100 contain crude-mixed phenols, sometimes referred to as "tar acids", and dissolved NH_3, H_2S and CO_2 gas as impurities.

The sour water is initially stripped to remove the dissolved gases, and the phenol containing water is treated with isopropyl ether for extraction of phenols.

The stripped gases (NH_3, H_2S, CO_2) are then treated using the Phosam-W process to recover ammonia. The ammonium phosphate solution absorbs 99% of the ammonia and the unabsorbed gases (H_2S, CO_2) are directed to the Claus sulfur plant in Section 600. Ammonia dissolved in the solution is recovered as an anhydrous byproduct and sent to storage (Section 800).

3.1.8 Storage and Shipping (Section 800) *

Facilities for receiving and storage of raw materials and consumable chemicals such as coal, diethanolamine (DEA), isopropyl ether, caustic, etc. have been provided for a 30-day production run based on full plant capacity.

Coal will be stored as open stock-piles near the coal preparation area. Water supplied to the plant by pipelines will be stored in a reservoir for 1-day maximum plant usage. This reservoir protects against a brief interruption of the raw water supply and also provides reserves for fire protection services.

All chemicals except caustic and DEA have been assumed to be delivered to the plant site in drums or bags and are stored in a warehouse. Caustic will be received by tankers (concentration 50°Be caustic), diluted to 20°Be during unloading, and stored in offsite tankage. A similar system has been provided for DEA. It will be received by truck as 100 wt% DEA, diluted to 30 wt% solution and stored.

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* not shown in block diagram

For intermediate products such as hydrocrackate (hydrogenated extract), raw naphtha, mid-distillate, and heavy oil (recycle solvent), storage capacity for 2-days of production has been assumed.

For byproducts such as phenols, ammonia, and sulfur, storage capacity sufficient for a 30-day production run has been considered. Ammonia will be stored in refrigerated spheres. Sulfur will be stored in a heated pit and shipped in molten form at 130 to 145°C.

For propane/butane (LPG), gasoline, jet fuel and diesel fuel, storage capacity for 5 days has been considered.

3.1.9 Utilities (Section 900)

The utilities service includes a full range of systems necessary for the operation of the plant, such as raw water treatment, cooling water and boiler feed water, condensate treatment, steam and electricity generation, plant and instrument air system, fuel gas purification and distribution.

A simplified diagram showing the various processing areas for water treatment is given in Figure 3-1. River water is assumed to be the source of make-up water for the plant. This water will be treated in a cold lime unit and filtered to reduce hardness, carbonate alkalinity, turbidity and suspended organics. The softened water is used directly as cooling tower make-up water and as process water. It is also supplied to the demineralization unit where cation, anion, and mixed-bed

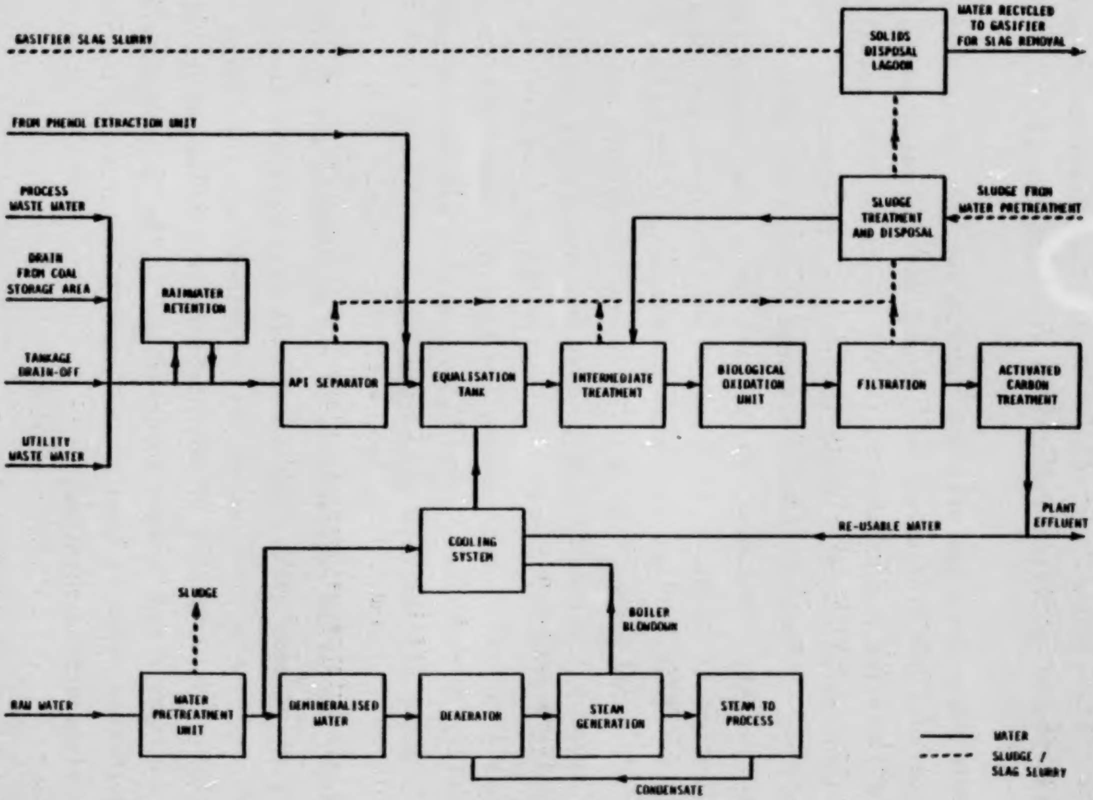

Figure 3-1 Simplified Flow Diagram of the Water Treatment Operation in the LSE Process

ion-exchangers are used for boiler feed water (BFW) preparation. The demineralized water is sent for deaeration and then distributed to onsite and offsite steam generators.

Waste water from the effluent control section, drained-off water from coal and product storage areas, and contaminated rain water are collected and treated in an API separator for removal of oil contamination before passing to the intermediate treatment section. Cooling tower blowdown water is also sent for intermediate treatment where the water is neutralized and subjected to chemical floculation and dissolved air flotation. The water is further subjected to biological oxidation followed by activated carbon treatment. Part of the treated water is used as make-up for the cooling towers, and the excess passes to the outfall system. Uncontaminated rainwater is also drained to the outfall system. Sludge collection from the waste-water treatment facility is thickened, aerobically digested and then concentrated on gravity belt filters before disposal.

Most of the steam requirements of the process is met by the recovery of process heat. The additional high-pressure steam required for power generation is produced by coal-fired steam boilers. Superheating of the high-pressure steam is provided by a fuel-gas-fired steam superheater. The high-pressure and superheated steam is used to generate electricity for the process via steam-driven turbines.

Fuel gases produced in Sections 300, 400 and 1100 are compressed and sent to fire-heaters via the fuel gas distribution system. Nitrogen from the air separation unit is distributed for inert gas blanketing, solid conveying, etc..

3.1.10 Offsite Facilities (Section 1000) *

Offsite requirements include items such as administration buildings, laboratory, maintenance equipment and buildings, warehouses, a fire-house, roads and onsite railway trackage.

3.1.11 Product Upgrading (Section 1100)

The primary distillate products, naphtha (C_5-200°C) and mid-distillate (200 - 300°C) are upgraded to produce transportation fuels such as gasoline, jet fuel and diesel.

The raw naphtha is catalytically hydrotreated for removal of heteroatoms and then reformed to produce a blendstock for gasoline. A part of the hydrogen required is supplied from Section 400, but most of the requirements are met by recovering hydrogen from the reformer gases. The stabilized reformate has a Research Octane Number of 108 and is blended with butane for volatility adjustment and with raw naphtha (RON 80) to yield unleaded gasoline.

The mid-distillate is also hydrotreated to yield jet fuel and diesel. The jet fuel meets most of the requirement of the Jet A specification (ASTM D1655-78). The diesel fuel has a Cetane number of 48 and meets the ignition requirements for most grades of diesel.

A detailed description of the product upgrading section with flow diagrams is given in Chapter 3.4.

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* Not shown in overall block diagrams

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

3.2 OVERALL MATERIAL BALANCE, PRODUCT EFFICIENCIES, AND UTILITIES SUMMARY

The interrelationship between the various processing areas and the overall material balance of a conceptual, commercial, integrated, self-sufficient LSE plant are presented in Drawings 3-1 and 3-2 based on 65% (MAF basis) and 82% (MAF basis) extraction efficiencies, respectively. Product yield and thermal efficiencies are summarized in Table 3-2.

For the 65% extraction case, about 14,152 t/d of MF Willow Bunch lignite is fed into the coal extraction reactors. A portion of the remaining coal (1,267 t/d MF Coal) is gasified along with residual char (7,333 t/d of coal liquefaction residues*) to produce the make-up hydrogen for the process. The balance (831 t/d MF coal) is used as a supplemental fuel for steam and power generation for a self-sufficient operation.

For the 82% extraction case, about 11,640 t/d MF coal is fed to the extraction reactors, 3,926 t/d MF coal is used in the gasifiers with 4,385 t/d of residual char** to produce make-up hydrogen for the process, and the balance, 684 t/d MF coal, is used as a supplemental fuel in the steam and power generation section.

The main products are gasoline, diesel and jet fuel totalling about 27,760 barrels and 29,320 barrels respectively for the 65% and 82% extraction cases. On MAF basis, the total liquid product corresponds to 2.48 bbl/t and 3.19 bbl/t, respectively, reflecting the different extraction levels in the reactor. However, the liquid product yield based on the total coal feed to the plant is nearly the same for both cases. The slightly lower yield in the 65% extraction case is due to the increased processing in the coal liquefaction section. The nearly equal liquid product yields for both extraction cases result from the fact that most of the char is efficiently separated by filtration without any significant loss of net liquid product and that all the residual char is completely utilized for the production of make-up hydrogen for the process.

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* includes 3914 t/d MAF char, 2970 t/d ash, and 449 t/d of hydrocarbon liquids

** includes 1656 t/d MAF char, 2441 t/d ash, and 288 t/d of hydrocarbon liquids.

The "overall product" efficiencies are 57.5% and 60.7% for the 65% and 82% extraction cases respectively. The definition of overall product efficiency is the total gross calorific value (GCV) of all products divided by the input energy of coal and electricity. In the present conceptual design, all the electricity required for the process is generated onsite, therefore, this term is zero. Similarly, the "liquid product" efficiencies pertaining only to gasoline, diesel and jet fuel, is 51% and 54% respectively.

For the 65% extraction case, a detailed material balance for each of the main processing units/areas is presented in Appendix E. The utility balance, i.e. consumption and production of steam, boiler feed water, condensate etc. is summarized in Table 3-3. The plant is self-sufficient in terms of steam and power. About 210 MW of electricity is required which is generated on site. Most of the steam is produced by recovering process heat and the balance is met by burning coal in coal-fired boilers.

Table 3-2 Product Yield and Efficiencies of the LSE Conceptual Plant

(Basis: 25,000 t/d of as-received coal)

Coal	Willow Bunch Lignite	
Extraction Efficiency, MAF basis	65%	82%
Coal (as-received)		
Moisture, wt%	35.0	35.0
Ash, wt% of MF coal	21.0	21.0
Moisture Content, wt%		
Coal fed to Reactor	2.0	2.0
Coal fed to Gasifiers*	5 to 8	5 to 8
Coal to Utilities **	28.5	28.5
Input, t/d		
MF Coal to Reactor	14,152	11,640
MF Coal to Gasifiers*	1,267	3,926
MF Coal to Utilities**	831	684
Electricity, MW	-	-
Output, t/d (bbl/d)		
Propane/Butane	389	410
Gasoline	1,490 (10,296)	1,572 (10,863)
Jet Fuel	1,402 (9,365)	1,481 (9,893)
Diesel Fuel	1,262 (8,102)	1,334 (8,564)
Sulfur	125	125
Ammonia	64	65
Phenol	32	32
Liquid Product Yield		
bbl/t MF coal to reactor	1.96	2.52
bbl/t MAF coal to reactor	2.48	3.19
bbl/t MF coal to plant	1.71	1.80
bbl/t MAF coal to plant	2.16	2.27
Product Efficiency, %		
Overall	57.5	60.7
Liquid	51.0	54.0

* Coal for make-up hydrogen and fuel gas production

** Coal for stream and power generation.

Table 3-3 Summary of Utility Balance for a Conceptual, Self-Sufficient, LSE Plant
(Coal Feed Rate: 25,000 t/d of as-received Willow Bunch lignite to the plant)
(Coal Conversion in Reactor: 65% MAF Basis)

CONSUMPTION

Section	Description	Electricity kW		Fuel Gas	Steam t/d						Water t/d		Air	Inert N_2
		Running	In-stalled	MJ/s	HP	MP	LP	Conden-sate	BFW	Boiler Blow-down	Process	Cooling	t/d	t/d
100	Coal Preparation	6977	11200	365	–	–	–	–	–	–	–	–	3	–
200	Coal Slurrying and Solvent Storage	2670	3610	–	–	–	–	–	–	–	–	–	3	–
	Slurry Preheating	–	–	30	–	–	–	–	1428	–	–	–	1.7	–
	Coal Extraction	286	360	–	–	–	–	–	–	–	–	38215	3.5	–
	Residue Separation	970	1700	–	–	–	30	–	–	–	–	37113	3.6	70
	Gas Treatment and H_2 Recovery	–	–	–	–	–	1750	–	–	–	–	17325	2.0	–
	Hydrocracking and Fractionation	68170	75850	180	–	–	–	–	6785	–	760	411990	13.5	–
300	Amine Plant and LPG Processing	25940	28820	37	–	8	252	–	–	–	–	47350	5.5	–
400	Gasification	3940	4380	–	–	796	–	–	8960	–	–	28365	22.0	600
	Fuel Gas Purification	–	–	–	–	–	–	–	–	–	–	–	–	–
	Shift Conversion	–	–	–	–	12384	–	–	5091	–	–	82920	5.5	–
	CO_2 Removal and Methanation	4155	4620	–	–	820	–	–	–	–	–	55280	5.5	670
500	Oxygen Plant	72800	76500	–	–	–	–	–	–	–	–	87950	6.0	–
600	Emission Control	160	175	4	–	–	–	–	406	–	–	2862	1.0	–
700	Effluent Control	1950	2605	–	–	–	415	–	–	–	82	1340	1.0	–
900	Steam and Power Generation	–	–	66	12025	–	–	–	15048	–	–	486710	3.0	–
900/1000	Auxiliary Facilities	7600	9490	–	–	792*	–	–	–	–	–	–	–	96
1100	Product Upgrading	16045	20050	23	–	453	126	–	–	–	223	114680	2.0	–
	Total Consumption	211663	239360	705	12025	15253	2573	–	37718	–	1065	1412100	81.8	1436

* Excess MP steam in auxiliary facilities appears as condensate.

Notes:

Make-up BFW = BFW (Consumed) - Recycle Condensate (Generated) - Recycle Boiler Blowdown (Generated) = 37718 -21554 - 2954 = 13210 t/d
Make-up BFW = Net Steam (Consumed) = 30 + 796 + 12384 = 13210 t/d
Make-up Cooling Water = Cooling Water (Consumed) - Cooling Water (Generated) = 1412100 - 1369129 = 42971 t/d
Make-up Process Water = Process Water (Consumed) - Process Water (Generated) = 1065 - 0 = 1065
Total Make-up Water = Make-up BFW + Make-up Cooling Water + Make-up Process Water = 13210 + 42971 + 1065 = 52746 t/d

Table 3-3 Summary of the Utility Balance for a Conceptual, Self-Sufficient LSE Plant (cont'd)
(Coal Feed Rate: 25,000 t/d of as-received Willow Bunch lignite to the plant)
(Coal Conversion in Reactor: 65% MAF basis)

PRODUCTION

Section	Description	Electricity kW		Fuel Gas	Steam t/d						Water t/d		Air	Inert N_2
		Running	In-stalled	MJ/s	HP	MP	LP	Conden-sate	BFW	Boiler Blow-down	Process	Cooling	t/d	t/d
100	Coal Preparation	-	-	-	-	-	-	-	-	-	-	-	-	-
200	Coal Slurrying and Solvent Storage	-	-	-	-	-	-	-	-	-	-	-	-	-
	Slurry Preheating	-	-	-	-	-	1400	-	-	28	-	-	-	-
	Coal Extraction	-	-	-	-	-	-	-	-	-	-	-	-	-
	Residue Separation	-	-	-	-	-	-	-	-	-	-	-	-	-
	Gas Treatment and H_2 Recovery	675	840	-	-	-	-	1750	-	-	-	-	-	-
	Hydrocracking and Fractionation	-	-	-	-	6656	-	-	-	129	-	-	-	-
300	Amine Plant and LPG Processing	-	-	450	-	-	-	260	-	-	-	-	-	-
400	Gasification	-	-	-	8100	-	-	-	-	860	-	-	-	-
	Fuel Gas Purification	-	-	255	-	-	-	-	-	-	-	-	-	-
	Shift Conversion	-	-	-	3925	-	706	-	-	460	-	-	-	-
	CO_2 Removal and Methanation	-	-	-	-	-	-	820	-	-	-	-	-	-
500	Oxygen Plant	-	-	-	-	-	-	-	-	-	-	-	-	1436
600	Emission Control	-	-	-	-	214	155	-	-	37	-	-	-	-
700	Effluent Control	-	-	-	-	-	-	415	-	-	-	-	-	-
900	Steam and Power Generation	210988	238520	-	-	8383	312	16938	-	1440	-	-	-	-
900/1000	Auxiliary Facilities	-	-	-	-	-	-	792*	-	-	-	1369129	81.8	-
1100	Product Upgrading	-	-	-	-	-	-	579	-	-	-	-	-	-
	Total Production/Recycle/Make-up	211663	239360	705	12025	15253	2573	21554	-	2954	-	1369129	81.8	1436

* Excess MP steam in auxillary facility appears as condensate.

Notes:

Make-up BFW = BFW (Consumed) - Recycle Condensate (Generated) - Recycle Boiler Blowdown (Generated) = 37718 - 21554 - 2954 = 13210 t/d

Make-up BFW = Net Steam (Consumed) = 30 + 796 + 12384 = 13210 t/d

Make-up Cooling Water = Cooling Water (Consumed) - Cooling Water (Generated) = 1412100 - 1369129 = 42971 t/d

Make-up Process Water = Process Water (Consumed) - Process Water (Generated) = 1065 - 0 = 1065

Total Make-up Water = Make-up BFW + Make-up Cooling Water + Make-up Process Water = 13210 + 42971 + 1065 = 52746 t/d

3.2 OVERALL MATERIAL BALANCE, PRODUCT EFFICIENCIES, AND UTILITIES SUMMARY (cont'd)

The make-up water requirement is about 57,246 t/d (10,500 U.S. gpm). Of the total requirement, 13,210 t/d is used as make-up BFW, 1,065 t/d as process water, and the balance, 42,971 t/d, is for cooling tower make-up and other losses. The large volume of this make-up water requirement is mainly due to excessive losses and the consumption of water in the water-cooling system. If air cooling is employed wherever possible (instead of the water cooling system as assumed in the conceptual design), the make-up water requirement would be reduced significantly. According to the estimate of the process developer, the total make-up water requirement would be less than 27,000 t/d (5,000 U.S. gpm), but the capital investment would increase by about 5.0 million dollars.

3.3 COAL LIQUEFACTION SECTION OF THE LIQUID SOLVENT EXTRACTION (LSE) PROCESS - SECTION 200

The coal liquefaction section of the Liquid Solvent Extraction (LSE) process may be subdivided into the following processing areas:

- Coal/oil slurry preparation, slurry drying and preheating, and extraction of coal.
- Solid/liquid separation and filter cake drying for solvent recovery.
- Hydrocracking (hydrogenation of extract), gas/liquid separation, gas treatment and hydrogen recovery.
- Liquid product recovery and fractionation.

A process flow diagram of the coal liquefaction section is presented in Drawing 3-3. A detailed description of the process flow diagram is given below.

3.3.1 Coal/Oil Slurry Preparation, Slurry Drying and Preheating, and Extration of Coal

Partially dried and pulverized coal is mixed with recycle oil (solvent) and the resulting slurry is pressurized, preheated and fed into the coal extraction reactor. Residual moisture present in the coal is removed by slurry drying.

Details:

Partially dried and pulverized coal (70% minus 0.075 mm, 15 wt% moisture) from section 100 is continuously fed into the atmospheric slurry-mix-tank/slurry-drier V-2 where simultaneous slurrying and drying of the coal is carried out. The pulverized coal is fed by means of rotary air locks to vessel V-2. Inert nitrogen gas is injected into the rotary air lock feeders to minimize the amount of atmospheric air entering vessel V-2. The inert gas, therefore, keeps the oxygen concentration inside the system below the explosive limits and also minimizes the loss of hydrocarbons with the vent gases.

For slurry preparation, hot recycle oil (solvent) is mixed with the feed coal in an oil-to-coal ratio of 2.0. Mixing is achieved by internal agitation resulting from recycling a portion of the coal/oil slurry via an external pump P-2. The slurry preparation step is carried out at about 135°C so as to reduce the moisture content of the feed coal from 15 wt% to the desired level of less than 2 wt%. Vapors leaving vessel V-2 are cooled in air-exchanger E-1 and in trim cooler E-2, and the effluents are collected in the slurry-drier overhead drum V-3. The condensed oil is recycled to vessel V-2 and the sour water is sent to the effluent control unit (Section 600) for further treatment. The gases are sent to flare.

The resulting coal/oil slurry from vessel V-2 is collected in slurry holding tank V-4 from where it is pumped to a pressure of 2 MPa by reciprocating pump P-5. Net positive suction pressure required for reciprocating pump P-5 is provided by centrifugal pump P-4. The pressurized slurry is then preheated to the extraction temperature of 400°C. Heating is initially achieved by heat exchange against the hot reactor-effluents in E-3 and then in fuel gas fired-heated FH-1. Waste heat from the combustion gases leaving FH-1 is recovered as low-pressure steam.

The pressurized and preheated slurry is fed into coal extraction reactor (digestor) EX-1 where the coal and solvent are contacted for a residence time of 60 minutes during which organic matter of the coal is dissolved. For Willow Bunch lignite, a typical extraction efficiency in the range of 65-82% MAF coal was considered. No catalyst or external hydrogen is used in extraction reactor EX-1.

The reactor effluents are cooled in exchanger E-3 and fed into hot separator V-5. Vapors leaving vessel V-5 are further cooled in exchangers E-4 and E-5 to condense water and hydrocarbon liquids. The uncondensed gases are sent for gas treatment (Section 300) and the sour water is sent to the effluent control unit (Section 700).

The slurry leaving hot separator V-5 is sent for solid/liquid separation.

3.3.2 Solid/Liquid Separation and Solvent Recovery

The coal-derived slurry is batch-filtered in order to separate the unreacted coal and ash (solid residues) from the liquid (extract and solvent). The filtered liquid is sent to catalytic hydrocracker.

The remaining liquid product present in the wet filter-cake is recovered by solvent wash followed by vaporization under vacuum using a steam ejector system. The dry filter-cake (solid residues) is sent to the make-up hydrogen and fuel gas production section.

3.3 COAL LIQUEFACTION SECTION OF THE LIQUID SOLVENT EXTRACTION (LSE) PROCESS - SECTION 200 (cont'd)

Details

The coal-derived slurry from hot separator V-5 is reduced in pressure through let-down valves and fed into batch-operated leaf filters for removal of solid residues. The filtration unit consists of four parallel trains because of equipment size limitation, each train consisting of three filters in operation and one spare. Each of these filters is equipped to carry out a complete filtration cycle but for simpler presentation purposes, only the technological lines in operation for a given filter are shown in the flow diagram. The three filters (FT-1a, FT-1b, and FT-1c) carry out in sequence the following three operations:

a) The coal-derived slurry is filtered to produce essentially a solid-free liquid stream and a wet filter-cake. Initial filtrate unsuitable for feeding to the hydrocracker is recycled to the filter.

b) The wet filter-cake is washed with a stream of recycle solvent and the washings are returned to the slurry preparation unit.

c) Most of the remaining solvent is recovered from the filter-cake by vaporization under reduced pressure using a 3-stage ejector system. The vapors leaving filter FT-1c are cooled in exchanger E-6 and the condensate is collected in vessel V-7 from where it is pumped to recycle-solvent drum V-1. The dry filter-cake (residues) containing less than 6 wt% liquid is mechanically removed from the filter and conveyed under a nitrogen blanket to the feed-hoppers of the gasifier unit for make-up hydrogen production.

3.3 COAL LIQUEFACTION SECTION OF THE LIQUID SOLVENT EXTRACTION (LSE) PROCESS - SECTION 200 (cont'd)

3.3.3 Hydrocracking of Extract, Gas/Liquid Separation, Gas Treatment and Hydrogen Recovery

The coal-derived extract from the filtration unit is fed into a catalytic hydrocracker operating at 420°C and 20 MPa. The effluents leaving the hydrocracker undergo a sequence of cooling and flash separation stages to condense and separate the hydrogenated extract from the gases.

High and medium-pressure gas streams are treated in amine absorbers and then 10% of the treated stream is cryogenically separated into a hydrogen-rich stream and a light hydrocarbon gas stream. Recovered hydrogen from the cryogenic unit is recycled to the hydrocracker and the sweet hydrocarbon gas stream is sent for light-ends separation (Section 300).

Low-pressure gases are sent directly to Section 300 for further processing.

Details

Filtered coal-derived liquid (extract and solvent) from the filtration unit is collected in vessel V-8. The combined liquid-stream is pressurized to 20 MPa by reciprocating pump P-8 and then mixed with a compressed hydrogen-rich gas stream. The resulting mixture is preheated to 420°C in hydrocracker-feed/exchanger E-7 and in gas-fired heater FH-2 before feeding into the fixed multi-bed hydrocracker reactors HC-1. Quench gas (cool hydrogen-rich gas) is injected between the bed to maintain the desired temperature by absorbing the exothermic heat of reaction.

3.3 COAL LIQUEFACTION SECTION OF THE LIQUID SOLVENT EXTRACTION (LSE) PROCESS - SECTION 200 (cont'd)

The hydrocracker effluent is cooled to 300°C in a series of heat exchangers (E-7, E-8, E-9) and fed into high-pressure flash drum V-9. Vapors leaving V-9 are further cooled to 40°C in exchangers E-10 and E-11. Before cooling in E-11, wash-water is injected into the vapor stream to prevent deposition of ammonium salts inside the exchanger tubes. The cooled effluent is collected in high-pressure drum V-10 where a separation between gases, liquid hydrocarbons and the aqueous phase is effected. High-pressure gases being rich in hydrogen are treated in amine absorber A-1 before being fed into the cryogenic unit CU-1 for recovery of hydrogen. Di-ethanolamine (DEA) is used as a solvent in absorber A-1. The liquid hydrocarbon separated from the sour water by decantation are reduced in pressure and sent to low-pressure drum V-14. The sour water is sent to the effluent control unit in Section 700.

The condensate leaving high-pressure flash drum V-9 is passed through a let-down valve and flashed in medium-pressure flash drum V-11 to recover the remaining hydrogen. Vapors leaving V-11 are cooled to 40°C in E-12 and collected in V-12 where a separation between gases, liquid hydrocarbons and sour water is effected. Medium-pressure gases leaving V-12 are compressed to 7.5 MPa, combined with high-pressure gases, and treated in amine absorber A-1 before hydrogen recovery.

Similarly, the condensate leaving the medium-pressure flash drum is flashed in low-pressure drum V-13. Vapors flashed off in V-13 are cooled in E-13 and then combined with the condensate from V-10 before final gas/liquid separation in V-14. Low-pressure gases leaving V-14 are combined with gases from vessel V-6 and sent for light-ends recovery (Section 300).

3.3 COAL LIQUEFACTION SECTION OF THE LIQUID SOLVENT EXTRACTION (LSE) PROCESS - SECTION 200 (cont'd)

The treated gas from the amine absorber A-1 is dried in molecular-sieve dryer D-1 to reduce the moisture content to less than 1 ppm before being fed into cryogenic unit CU-1. In CU-1, the gas stream is successively cooled to minus 185°C in two stages in plate-fin exchangers. In the first stage, the condensed liquid containing mainly propane and butane are separated, whereas in the second stage, methane and ethane are separated. The uncondensed hydrogen-rich gas stream leaving the cryogenic unit is preheated by incoming feed, and then pressurized to 20 MPa before being recycled the hydrocracker HC-1.

Refrigeration for the cryogenic unit is provided by adiabatic expansion of condensed liquids across the Joule-Thompson valves. The expanded gases (C_1-C_4) are sent for LPG processing to Section 300.

3.3.4 Liquid Product Recovery and Fractionation

The condensed liquid hydrocarbons obtained after gas/liquid separation of hydrocracker effluents are fed into a debutanizer for removal of lighter hydrocarbons ($< C_4$). The debutanizer bottoms are preheated and fed into a fractionator to yield primary distillate products: naphtha (C_5-200°C), mid-distillate (200-300°C) and heavy oil ($>$ 300°C).

Details

The liquid hydrocarbon condensates from vessels V-6, V-12, V-13, and V-14 are combined and fed into debutanizer DB-1. The light hydrocarbon gases leaving the debutanizer contain acid-gas impurities and are sent for gas treatment in Section 300.

3.3 COAL LIQUEFECTION SECTION OF THE LIQUID SOLVENT EXTRACTION (LSE) PROCESS - SECTION 200 (cont'd)

The debutanizer bottoms are preheated in exchangers E-19, E-20 and fired heater FH-4 before being fed into fractionator F-1. Hot combustion gas leaving the gas-fired heater is used to generate low-pressure superheated steam.

In the fractionator, low-pressure steam is introduced to reduce the flash point of the bottom product.

The overhead product leaving fractionator F-1 is cooled in E-23 to condense and separate the raw naphtha from sour water. Raw naphtha (C_5-200°C) separated by decantation in vessels V-18 is sent to the product upgrading section for futher refining. The sour water is sent to the effluent control unit (Section 700) for treatment.

The middle stream from the fractionator is fed into steam-stripper ST-2 from where a side mid-distillate stream (200-300°C) is withdrawn, cooled in exchangers E-24 and E-25, and sent to the product upgrading section for further refining. The bottom product from steam stripper ST-2 is used as a wash solvent (300-370°C) for the filters.

The fractionator bottoms is heavy oil (>300°C) which is recycled to the coal/oil slurry preparation unit.

3.4 PRODUCT ANALYSIS AND UPGRADING SCHEMES

In the Liquid Solvent Extraction (LSE) process, the coal-derived extract is catalytically hydrocracked (hydrogenated) to yield distillate products (C_5-370°C). The heavy oil (> 300°C) is recycled as a solvent for the process. Naphtha and mid-distillate may be further upgraded to gasoline, jet fuel, and diesel. Typical analyses of the intermediate and final products obtained by NCB for other coals are presented along with the suggested upgrading schemes.

Typical analyses of the coal-derived extract and hydrocracked product obtained with cobalt-molybdate (COMOX 471) as catalyst are given in Table 3-4.

The results indicate that between 80 to 90% of the heteroatoms are removed in the catalytic hydrocracking operation. Typical analyses of naphtha, mid-distillate and heavy oil are summarized in Table 3-5. The results indicate that the coal-derived primary distillate products have a low-sulfur content (< 0.1 wt%), relatively higher oxygen and nitrogen contents (0.1 to 0.4 wt%), and the products are highly aromatic in nature. Sulfur and nitrogen are mainly present in endocyclic groups and oxygen is present in phenolic groups.

The primary products do not conform to standard specifications for gravity, viscosity, smoke point, Cetane Number, Research Octane Number, etc. Consequently, further processing and upgrading steps are required to convert them into marketable products.

Table 3-4 Typical Analyses of Coal-Derived Extract and Hydrocracked Product
(U.K. Bituminous Coal, CRC-702)

	Coal-Derived Extract	Hydrocracked Product
Ultimate Analysis wt%		
Carbon	82	88
Hydrogen	7	11
Oxygen	9	1.0
Nitrogen	1	0.2
Sulfur	1	< 0.1
Removal of Heteroatoms, %		
Oxygen	-	90
Nitrogen	-	80
Sulfur	-	90
Distillation Result, Cummulative wt%		
C_5-170°C	0	22
170-250°C	0	40
250-350°C	7	70
Structural Types in Distillates ($<$ 350°C), wt%		
Paraffins	-	9
Cycloparaffins	-	32
Hydroaromatics	-	26
Alkylenzenes	-	14
Diaromatics	-	15
Triaromatics	-	4

Sources: NCB Report, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", Feb. 1981.

M.D. Gray, Coal Processing Consultants Ltd., "Liquids from Coal Processes Under Development by the National Coal Board, U.K." paper presented at 29th Canadian Chemical Engineering Conference on Synthetic Liquids from Coal and Biomass, held at Sarnia, Ontario, Sept. 30 - Oct. 3, 1979.

Table 3-5 Typical Analyses of Primary Coal-Derived Distillate Products

	Naphtha (C_5-200°C)	Mid-Distillate (200-300°C)	Heavy Oil* (>300°C)
Ultimate Analysis, wt%			
Carbon	87.0	88.7	90.5
Hydrogen	12.4	10.6	9.0
Oxygen	0.5	0.5	0.2
Nitrogen	0.2	0.2	0.2
Sulfur, ppm	50.0	50.0	1000.0
Hydrogen-to-Carbon Atomic Ratio	1.7	1.5	1.2
Group Types, wt%			
Paraffins	20	5	-
Naphthenes	40	40	-
Aromatics	40	55	-

* Heavy oil is the recycle oil and contains mainly polynuclear hydroaromatics. It is recycled to the slurry-preparation unit.

Sources: NCB Report, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", Feb. 1981.

M.D. Gray, Coal Processing Consultants Ltd., "Liquids from Coal Processes Under Development by the National Coal Board, U.K." paper presented at 29th Canadian Chemical Engineering Conference on Synthetic Liquids from Coal and Biomass, held at Sarnia, Ontario, Sept. 30 - Oct. 3, 1979.

3.4.1 Upgrading of Naphtha

The raw naphtha having a Research Octane Number (RON) of about 80 is initially hydrotreated mainly to remove sulfur and nitrogen-containing compounds which would otherwise poison the reformer's platinum catalyst. The naphtha hydrotreatment is very similar to that described in detail for mid-distillate fraction. (Chapter 3.4.2). Laboratory scale experiments on catalytic reforming over a commercial Pt/Al_2O_3 catalyst show that hydrotreated naphtha may be reformed to C_6-C_8 aromatic product with RON of about 108. The yield of reformate is about 90 wt% and contains 80 to 85 wt% of aromatic BTX fraction.

Figure 3-2 is a process flow diagram of the reforming section. Hydrotreated naphtha from intermediate storage tank V-1 is mixed with recycle hydrogen-rich gas and preheated in reactor feed/effluent exchanger E-1 and fired-heater FH-1. The reaction is carried out in three stages to allow reheating of the reactants in second and third stages.

To maintain constant catalyst activity in the reformer, spent catalyst is removed batchwise from the bottom of the third-stage reactor, regenerated, and recycled to the first-stage reator.

The effluent leaving the third stage of the reformer passes through exchanger E-1 and air cooler AE-1 into a series of two separators V-2 and V-3. In these vessels, hydrogen-rich gases are removed from the reformate and recycled to the naphtha and mid-distillate hydrotreaters. The reformate from V-3 is fed into a stabilizer column V-4 where light hydrocarbon gases ($< C_4$) are removed for use as fuel gas for the process after gas treatment.

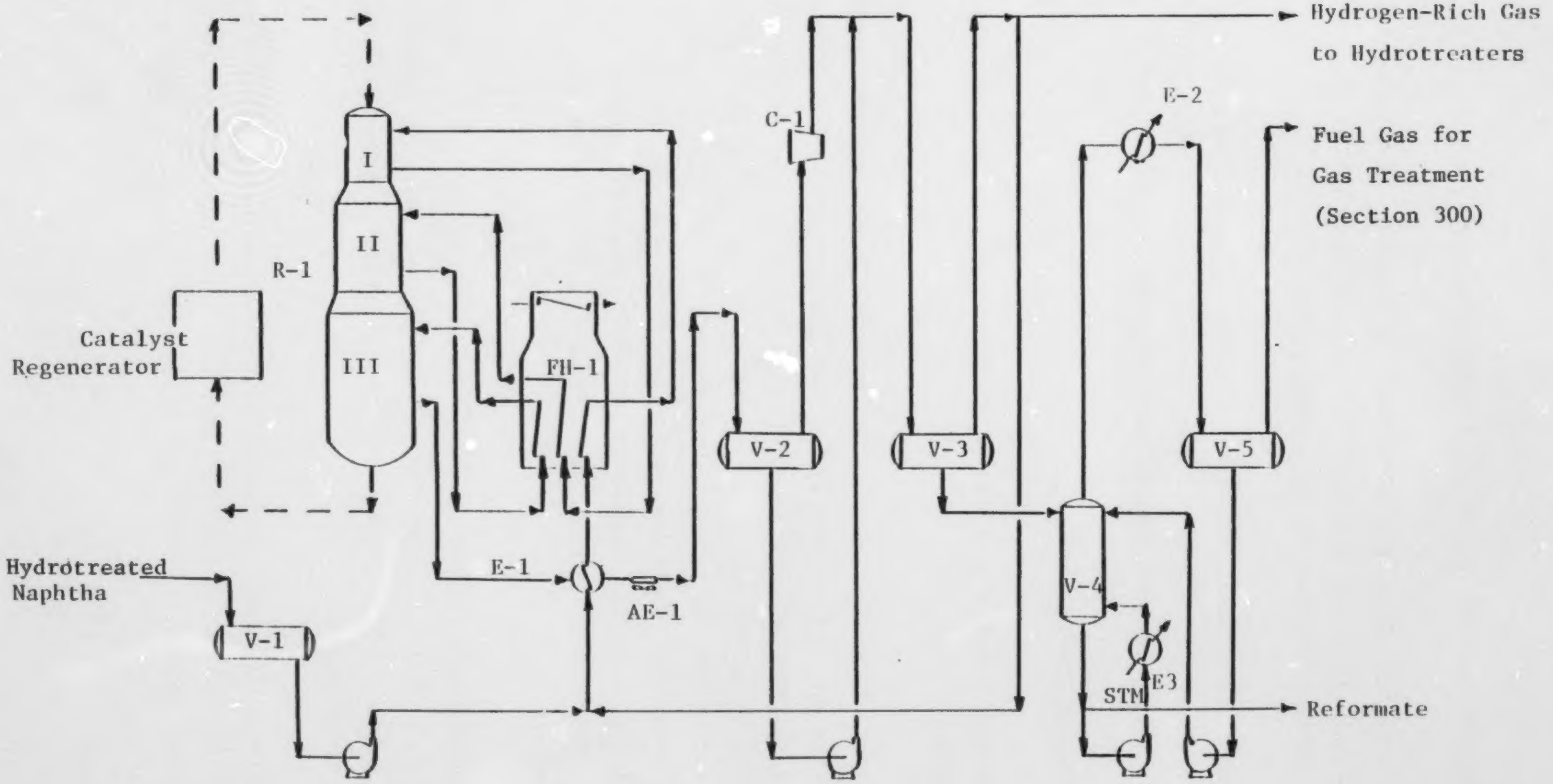

Figure 3-2 Process Flow Diagram of the Catalytic Reforming Operation of Hydrotreated Naphtha

Source: "Coal Liquefaction for Transport Fuels Using The Solvent Extraction Technology of U.K. National Coal Board", report prepared by NCB, February 1981

3.4 PRODUCT ANALYSES AND UPGRADING SCHEMES (cont'd)

The stabilized reformate (RON 108) is blended with butane for volatility adjustment and with raw naphtha (RON 80) to produce a final gasoline product having a RON of 98.

Pertinent properties of the unleaded gasoline product are summarized in Table 3-6. Comparison with gasoline of ASTM D493 standard show that the product meets most of the specifications.

3.4.2 Upgrading of Mid-Distillate

The raw mid-distillate stream is hydrotreated to produce jet fuel and diesel.

A process flow digram of the mid-distillate hydrotreatment section is given in Figure 3-3. Raw mid-distillate from intermediate storage tank V-1 is mixed with hydrogen, and the mixture is preheated in reactor feed/effluent heat exchangers (E-1, E-2) and fired-heater FH-1. The hot mixture is fed into catalytic hydrotreater HT-1 where the exothermic hydrogenation reaction is carried out. A side stream of quench hydrogen is injected to absorb the exothermic heat and to maintain the desired reaction temperature.

The hydrotreater effluent is cooled in exchangers E-1, E-2 and in air cooler AE-1. Quench water is injected into the process stream before cooling in AE-1 in order to dissolve ammonia formed in the hydrotreater.

The effluent is then flashed in high-pressure separator V-4 to recover the unreacted hydrogen gas, a portion of which is recycled to the hydrotreater and the balance is sent to the fuel gas systems.

Table 3-6 Properties of Lead-Free Gasoline Produced from Hydrotreatment and Reforming of Coal-Derived Naphtha

		ASTM D439
Density at 15.6°C kg/m^3	770	-
API Gravity	52	-
Reid Vapor Pressure, kPa	83	-
Distillation Analysis, °C		
IBP	32	-
10 vol%	65	55 - 65
50 vol%	109	76 - 115
90 vol%	183	185 max
FBP	208	225 max
Octane Rating		
Research Octane Number	98	87 min
Motor Octane Number	89	82 min
Antiknock Index $\frac{(RON + MON)}{2}$	93.5	93 - 95
Copper Strip Corrosion	2	1 max
Sulfur Content, ppm	0.5	1500 max
Lead Content	Nil	0.05 g/US gal

Sources: "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", report prepared by NCB, February 1981.

W.P. Syrett, NCB, Personnel Communication to J.A. Mikhlin, SNC Inc., NCB Reference No. DR 496/2, 7/3, Jan. 25, 1982

The hydrotreated liquid stream from high-pressure separator V-4 is further flashed in low-pressure separator V-5 to remove any residual hydrogen and lighter hydrocarbon gases which are sent to the fuel gas system. Sour water is sent to the effluent control system.

The hydrotreated liquid product from V-5 is sent to steam stripper ST-1 to remove the residual light-hydrocarbon gases. A vapor stream leaving ST-1 is cooled in AE-2 to condense and separate the light fraction (jet fuel) from the hydrocarbon gases in vessel V-6. Sour gases from V-6 are sent to the gas purification in Section 300 and sour water is sent to the effluent control section for water treatment.

Another hydrotreated liquid product stream from ST-1 is fed into a vacuum column to remove traces of moisture present in the stream. The final hydrotreated product leaving VT-1 is cooled in E-5 and sent to diesel fuel storage tanks.

Table 3-7 presents the pertinent properties of the jet fuel and diesel in comparison with ASTM standard specifications. The jet fuel has a higher density than the specification but it may be controlled by adjusting the jet-fuel cut. The diesel fuel meets all the pertinent ASTM specifications.

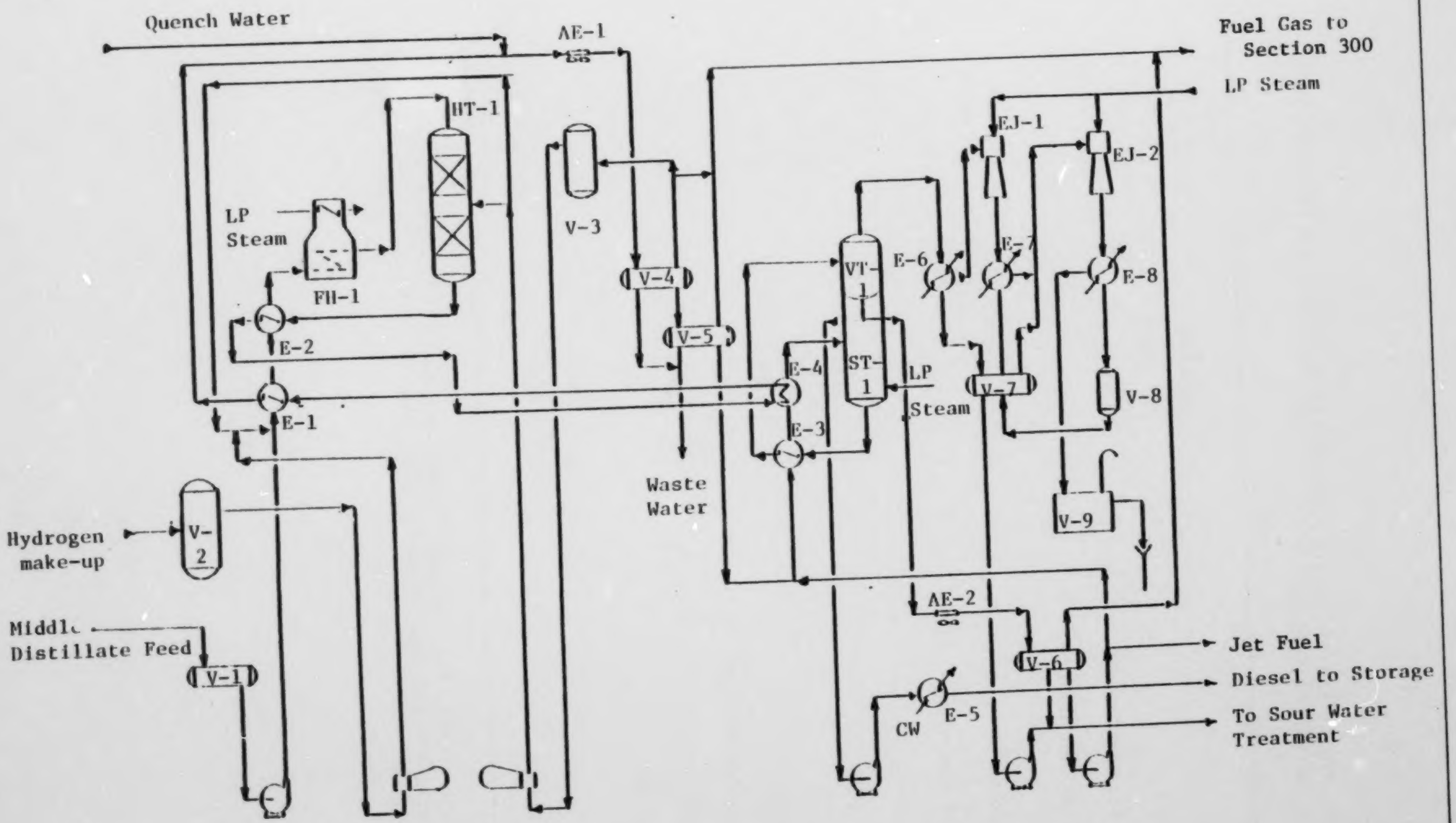

Figure 3-3 Process Flow Diagram of Hydrotreatment Operation of Mid-Distillate

Source: "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of UK National Coal Board", report prepared by NCB, February 1981

Table 3-7 Properties of Jet Fuel and Diesel Produced from Hydrotreatment of Mid-Distillate

	Jet Fuel	ASTM D-1655
API Gravity	33.8	37 - 51
Specific Gravity, g/cm^3	0.856	0.7753 - 0.8398
Distillation Temperature, °C		
10%	184	205 max.
End Point	199	311 max.
Net Heat of Combustion, KJ/kg	43682	42745

	Diesel Fuel	ASTM D-975
Cetane Number	48	40 min
Distillation Temperature, °C		
90 vol%	277	287 max
Flash Point, °C	92	38 min.
Ash, wt%	Nil	0.01 max
Sulfur, ppm	180	5000

Source: W.P. Syrett, NCB, Personal Communication to J.A. Mikhlin, SNC Inc., NCB Ref. No. DR/496/2, 7/3, Jan. 25, 1982.

3.5 ENVIRONMENTAL ASPECTS

Various solid, liquid and gaseous wastes streams are generated in the conceptual, commercial LSE plant. The environmental discharge of these wastes will have to be in compliance with the regulatory environment control acts of the province of Saskatchewan.

At present, there are five environment control acts covering air pollution, water resources management, water rights, ground water conservation and public health. A summary of provisions and requirements under these acts is presented in Appendix D.

The environment aspects pertaining to the LSE plant processing Willow Bunch lignite are outlined in the following pages.

3.5.1 Solid Wastes

Large amount of solid waste will be generated in the lignite preparation and gasification area. This waste will mainly consist of slag from gasifier bottom (about 3250 t/d), ash from offsite boilers (about 150 t/d) and fly ash from the electrostatic precipitators. The principal constituents of the solid waste would be the mineral composition of Willow Bunch lignite. The gasifier slag is granulated and extracted in a water lagoon before its disposal to mine site. This will reduce the possibility of leaching of trace metals from the slag into ground or surface waters. Ash from offsite boilers and fly ash will be disposed of as a slurry cake through the open ash lagoon system. The system will require a water make-up supply due to evaporation losses from the lagoon and water trapped in the lagooned ash. Since good quality water is not required for this service, cooling water blowdown may be used.

The dissolved solid content of the water will be high due to evaporation losses and ash leaching. Hence the lagoon will have a purge stream for the disposal of some of the supernatant matter. The ash disposal in this manner lowers the leaching of trace metals into ground or surface waters and therefore will be environmentally acceptable.

3.5.2 Liquid Wastes

Liquid waste will consist of sludge from the thickener of the raw gas wash-water system. Waste sludges will also be generated during the dimineralization of raw water and boiler feed water treatment. Another major contaminated stream will be sour water generated by the wet scrubbers where the raw gases produced by the gasifier are cleaned. A simplified flow diagram of sludge and waste water treatment, to meet environmental regulations, was presented earlier in Figure 3-1.

3.5.3 Gaseous Emissions

The proposed configuration of the indirect liquefaction plant includes an emission control section consisting of claus sulfur plant and SCOT tail gas treatment unit. The various gaseous emission streams from the entire plant are combined and treated in this unit which is designed to control the SO_2 emission within 100 ppm. This level of SO_2 discharge to atmosphere is environmentally acceptable under Saskatchewan's air quality standards (see Table D-1, Appendix D).

3.6 ECONOMIC ASSESSMENT OF LSE PROCESS

Preliminary capital cost estimate, operating and production costs of LSE process has been performed for a self-sufficient, integrated plant processing 25,000 t/d of Willow Bunch lignite. These costs have been developed in accordance with the criteria described earlier and following the procedure developed by SNC and approved by the Study Committee.

Product costs have been calculated for both operating cases (65% and 82% extraction efficiency) on "Energy Value" and "Product Value" bases. Product cost on "Energy Value" basis is calculated considering heating value only whereas the other also considers the market value and quality of the product. The latter approach is also known a ESCOE* methodology.

3.6.1 Financial Assumption and Economic Criteria

Some of the basic financial assumptions and criteria for economic evaluation of coal conversion processes are summarized in Table 3-8. Other data and economic criteria received from NCB and the Study Committee are presented later in the text.

The LSE process is evaluated as a self-sufficient, grass-roots facility designed to process 25,000 tonnes per day of as-received Willow Bunch lignite supplied to the plant. The plant is assumed to be located within 16 km (10 miles) of the coal mines in the Willow Bunch area, Saskatchewan. Coal, air and raw water are the major process inputs to the plant. All other utility requirements such as steam, fuel gas and electricity are generated on site.

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* K.A. Rogers an R.F. Hill, The Engineering Society Commission on Energy Inc., "Coal Conversion Comparison", DOE Report No. FE-2468-51, July 1979

3.6.1 **Financial Assumptions and Economic Criteria** (cont'd)

The feed coal to the plant is assumed to be sized to minus 5 cm and having typical (average) ultimate and proximate analysis as shown in Appendix C.

All costs are calculated in "constant" Canadian mid-1982 dollars and no forward escalation beyond mid-1982 included. Cost estimates and data received from process developer for a different plant location, capacity and conditions have been converted and adjusted according to standard industry index.

For the LSE plant, a 4-year construction period, 30-year operating life, straight-line depreciation and zero salvage value are assumed. The capital investment schedule during construction is based on a typical S-type curve (see Appendix G-1) and the full design-capacity level is expected to be achieved within 3 years of the start-up date (see Appendix G-2).

Cost of capital is assumed to be 10% per annum. For 100% debt financing, the value is equivalent to debt cost* in "real" terms. For 100% equity financing, it may be considered as return on equity before taxes.

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* For calculations in "constant" dollars, the debt cost is expressed in real terms, i.e. prevailing rates minus inflation.

Table 3-8   Financial Assumptions and Economic Criteria

Coal feed: 25,000 tonnes per day to plant.

Constant Canadian dollars, mid-1982.

Plant location: within 16 km (10 miles) of coal mines.

4-years construction period.

Capital investment schedule (S-type curve)

Year	%
- 4	9
- 3	25
- 2	36
- 1	30

Investment tax credit for cost of borrowing during construction claimed during year of occurence.

Plant life: 30 years

Straight-line depreciation method

Zero salvage value

Process contingency: 5%

Project and estimating contingency: 10%

Plant start-up efficiencies

Year	% Design Capacity
1	50
2	90
3	100

Plant service factor: 0.85 (310 operating days per year)

Cost of capital financing = 10% per annum

Land cost: $990/hectare ($400/acre)

Working capital: 3% of installed-plant cost

3.6.2 Procedure for Adjusting the Capital Cost Estimate and Total Investment to Saskatchewan

Capital cost estimates and investments data* available from the National Coal Board has been adjusted to the designated Saskatchewan area following the criteria and procedure described below. In the NCB data supplied to SNC Inc., Gulf Coast, U.S.A., was considered as a reference location.

- Equipment items are assumed to be 25% Canadian content and 75% U.S. content. Items purchased in U.S. are not available in Canada, otherwise a duty of 12.8% will be levied (Customs Canada, 27 Jan. 1982).

- All bulk materials (pipings, fittings, etc.,) are considered as 100% Canadian content, comprising 45% labour and 55% material components.

- The U.S. material and equipment content (75% of the total) is converted to Saskatchewan base by multiplying by a factor of 1.19 composed strictly of currency exchange rate (Royal Bank 27-01-82).

- The Canadian material and equipment content (25% of the total) is obtained by multiplying the cost given for a U.S. location by a shop-labour price differential factor of 1.1592 (see Appendix G-3). Including the exchange rate of 1.19, the "material plus currency differential factor" is 1.379.

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* W.P. Syrett, NCB report to SNC Inc., "The Processing of Saskatchewan Lignite by the NCB's Liquid Solvent Extraction Process: Willow Bunch Lignite", NCB Ref. No. DR 496/2, 7/3, January 23, 1982. The report is attached in confidential Volume V, "Process Developers Data and Information".

3.6.2 **Procedure for Adjusting the Capital Cost Estimate and Total Investment to Saskatchewan** (cont'd)

- Construction manpower/labor costs for a reference location are converted to Saskatchewan location by multiplying by a productivity adjustment factor of 1.42 (see Appendix G-4).
- Capital cost estimates are adjusted for inflation to the base year (mid-1982) using the Chemical and Mineral Plant Price Index given in Appendix G-5.

Following the above procedure, the capital cost estimate of LSE plant with breakdown by individual processing area/units is given in Table 3-9. All costs except those for the gasification, raw gas compression and acid gas removal (Section 400) is based on NCB data supplied to SNC Inc. NCB data in U.S. dollars (October 1981) for a U.S. Gulf Coast location has been converted to Canadian dollars (mid-1982) and Saskatchewan location according to the criteria and guidelines mentioned above. For Section 400, the costs reported in U.S. dollars (January 1982) by Krupp Koppers GmbH in their confidential report* to SNC Inc. has been adjusted for capacity difference, exchange rate, material and labor rate differential.

The total capital investment for a LSE plant is given in Table 3-10. The installed-plant cost includes all the equipment items, bulk materials, labor, engineering and support, process contingency, licenses and prepaid royalties, and other construction costs. A project and estimation contingency of 10% has been added to cover uncertainties in cost estimation, equipments damage during construction, delay in equipment delivery and project schedule.

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* H. Staege, Krupp Koppers GmbH report to SNC Inc., VK-1/Km-Mo-1813, June 1982. The report is attached in confidential Volume V, "Process Developers Data and Information".

Table 3-9 Capital Cost Estimate of LSE Plant, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of "as-received" Willow Bunch Lignite]

Section	Process Area/Unit	Equipment Items	Bulk Material and Labour	Engineering and Support	Installed Cost
100	Coal Preparation	23.29	20.37	5.90	49.56
200	Coal/Oil Slurry Preparation	2.20	6.36	1.67	10.23
	Slurry Preheating	2.36	4.36	1.34	8.06
	Coal Extraction	4.88	14.94	3.95	31.83
	Gas/Vapor/Slurry Separation and Hydrocracker Feed Preparation	41.24	77.45	24.97	143.66
	Hydrogen Recovery and Cryogenic Unit	3.25	3.19	0.70	7.14
	Residue Separation	28.61	43.56	13.70	85.87
	Hydrocracking	42.66	114.95	32.33	189.94
	Fractionation	9.00	18.39	7.02	34.41
300	Amine Plant	5.69	15.05	4.55	25.29
	Light-Ends Recovery and LPG Processing	12.98	32.19	10.0	55.17
400	Gasification	144.76	65.31	*	210.07
	Shift Conversion	43.45	18.25	*	61.70
	Acid Gas Removal and Methanation	115.78	61.41	*	177.19
500	Oxygen Plant	127.41	*	31.85	159.26

* Included in the equipment items.

Table 3-9 Capital Cost Estimate of LSE Plant, in Millions of Canadian Dollars (mid-1982) (cont'd)

[Capacity: 25,000 t/d of "as-received" Willow Bunch Lignite]

Section	Process Area/Unit	Equipment Items	Bulk Material and Labour	Engineering and Support	Installed Cost
600	Emission Control	2.37	4.47	1.52	8.36
700	Effluent Control (incl. Waste Water Treatment	80.58	*	16.03	96.61
800	Product Storage and Blending	**	**	**	**
900/1000	Steam and Power Generation	183.53	*	43.61	227.14
	Flue Gas Desulfurization	9.56	18.30	4.07	31.93
	Ancillary and Offsite Facilities	219.64	28.67	58.93	307.24
1100	Distillate Hydrotreatment	13.55	32.06	11.58	57.19
	Reforming	4.55	6.52	2.86	13.93
	Sub-Total	1121.34	585.8	276.60	1983.74
Adjustment Factor for Canadian and U.S. Content (0.25 x 1.1592 + 0.75 = 1.0398)		44.63	-	-	44.63
	Total	1165.97	585.8	276.60	2 028.37

* Included in equipment items.
** Included in ancillary and offsite facilities.

Sources: W.P. Syrett, NCB report to SNC Inc., DR-496/2, 7/3, January 23, 1982.

H. Staege, Krupp Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982.

3.6.2 **Prodedure for Adjusting the Capital Cost Estimate and Total Investment to Saskatchewan** (cont'd)

Initial catalysts, chemicals and operating supplies cover the cost of "initial charge" of these materials to the plant. These costs have been assumed to be equivalent to 1-month supply catalysts, chemicals and other consumables (except water) based on full plant-capacity.

Start-up and other costs taken as 3% of total plant installed cost (TPIC) to cover for operating expenses during commissioning and start-up period. These costs also include additional capital for modifications required to bring the plant to full capacity, associated process and environmental studies, contractor fees, owners engineering offices and cost for obtaining construction permits.

Working capital is assumed as 3% of TPIC. Land costs have been calculated for a 220-acre plot considering \$990 / hectare (\$400/acre). The size of the land has been estimated from the plot plan supplied by NCB.

Interest during construcion is based on capital spending schedule given in Table 3-8 and Appendix G-2. Investment credits equal to 48% of interest during construction has been claimed in the year of occurrence.

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1 acre = 4047 m^2

Table 3-10 Breakdown of Total Capital Investment for LSE Plant, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of "as-received" Willow Bunch lignite]

Installed-Plant Cost (including process contingencies)	2028.37
Project and Estimation Contingencies	202.84
Total Plant Installation Cost (TPIC)	2231.21
Initial Catalysts, Chemicals and Operating Supplies	1.07
Start-up and Other Costs	66.94
Working Capital	66.94
Land	0.09
Interest During Construction	272.75
Total Capital Investment, "C"	2639.00

Source:

W.P. Syprett, NCB report to SNC Inc., DR-496/2, 7/3, January 23, 1982
M.C. McBeath, ESCOE, DOE Report FE-2468-44, April 1979

3.6.2 Procedure for Adjusting the Capital Cost Estimate and Total Investment to Saskatchewan (cont'd)

The total investment of the LSE process is Canadian $2,639 million for 25,000-t/d self-sufficient, integrated, commercial plant including all the processing units, utilities and ancillary facilities. The total investment, however, does not include:

- The cost of coal mine, mine development and transportation facilities from the mine to the plant.
- Railways, roads and other transportation facilities outside the plant site.
- Product distribution pipelines.
- Water supply and pipelines outside the plant site.
- Housing, community facilities and associated infrastructure.

3.6.3 Operating, Maintenance and Production Costs

Operating, maintenance and production costs for a LSE plant for two cases (65% and 82% reactor extraction efficiency on MAF basis) is presented in Table 3-11. These costs are reported on an annual basis assuming a plant service factor of 85%, i.e. 310 operating days/year at 100% design capacity (25,000 t/d of coal).

Operating and supervisory labour costs is calculated using a typical manpower distribution and average salary scale given in Appendix G-6. The total manpower estimated by NCB for LSE plant is 900, consequently, the operating labour force is 306 (34% of 900). Operating supplies are estimated at 30% of operating labour cost. A labour burden totalling 35% of direct labour cost (operating and supervisory) is added to cover fringe benefits, sick leave, social security, unemployment insurance and shift differentials, etc.

Table 3-11 Annual Operating, Maintenance and Production Costs of LSE Plant, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of "as-received" Willow Bunch Lignite]
[310 operating days per year]

	65% extraction MAF basis	82% extraction MAF basis
Fixed Expenses		
Operating Labour*	6.88	6.88
Operating Supplies	2.06	2.06
Supervisory Labour*	1.89	1.89
Labour Burden	3.06	3.06
Maintenance	66.94	66.94
Property Taxes and Insurance	61.36	61.36
Plant Overhead	32.64	32.64
Variable Expenses		
Coal Feed	75.62	75.62
Catalysts and Chemicals	10.80	10.80
Make up Water	4.30	4.10
Waste Disposal	2.00	2.00
Running Royalties	-	-
Total Annual Operating Costs	267.55	267.35
Less Byproduct Credits	12.95	13.04
Total Annual Net-Operating Costs	254.60	254.31
Capital Costs (k = 0.11155)**	294.40	294.40
TOTAL	549.00	548.71

* Does not include maintenance labour and supervision as these items are covered in maintenance cost. Similarly, plant management and administrative staff are included in plant overhead.

** Value of capital cost factor "k", as defined in Equation 2, is 0.11155.

Sources:

W.P. Syrett, NCB report to SNC Inc., Ref. No. DR 496/2, 7/3, January 23, 1982.

B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979.

R. Salmon et al, Oak Ridge National Laboratory, Report No. ORNL/TM-7581, April 1981.

3.6.3 **Operating, Maintenance and Production Costs** (cont'd)

Maintenance is estimated at 3% of total plant installation cost (TPIC). This includes maintenance materials, replaced equipment items, maintenance labour, supervision, labour burden and other regular maintenance contracts.

Property taxes and insurance are assessed at 2.75% of the total plant installed cost (TPIC) comprising 2.0% local taxes and 0.75% insurance.

Plant overhead is assumed to be 40% of the operating and supervisory labour, maintenance and operating supplies. This cost includes plant management, engineering, fire protection, medical services, personnel, accounting, clerical and janitorial services.

Cost of coal, catalyst and chemicals, utilities and waste disposal is treated as a variable expense. Cost of coal (Willow Bunch lignite) supplied to the plant is $9.75/tonne. This price was provided by Saskoil* and includes $0.28/tonne transportation costs from the mine to the plant.

Raw water is the only utility purchased from an outside source and it is available to the plant site at a cost of $1.10 per 1000 imperial gallons**. This cost has been estimated by SNC-Keith-Gabriel, Saskatchewan, and includes cost of canal and water pumping costs from Diefenbaker Lake.

Waste disposal cost (for ash) is assumed to be $2/tonne.

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* F. Mourits, Coal Liquefaction Study Management, letter to J.A. Mikhlin, Dec. 17, 1981.

** 1 imp gallon = 4,546 litres

3.6.3 **Operating, Maintenance and Production Costs** (cont'd)

Running royalties are payment on an annual or semi-annual basis to the process licensor for the use of a proprietary process. The nature of royalty payment is complex and requires more detailed information than that possible within the scope of this assessment. Therefore, the running royalties has been neglected.

By product credits for sulfur at $65/t, NH_3 at $300/t and mixed phenols at $250/t have been taken into consideration. The credit for CO_2 byproduct has not been considered due to market uncertainty.

Levelised capital costs and investment related charges is calculated to be $294.4 million per annum in "constant" dollars as shown in Table 3-11. In the first and second year of operation, the plant is operating at 50% and 90% of design capacity. To account for deficits in these two years, bridge financing at 10% interest rate in "real" terms and amortized over the balance period (28 years) is assumed. From the third year of operation, the plant is expected to operate at 100% design capacity, i.e., processing 25,000 t/d of coal.

3.6.4 **Product Costs**

Product costs of LSE Plant for the two cases (65% and 82% extraction efficiency) are presented on "Energy Value" and "Product Value" bases. Product cost on "Energy Value" basis is calculated considering heating value only, whereas the "Product Value" also takes into account the relative market value of individual products. The product costs on both "Energy Value" and "Product Value" bases are meant for relative ranking and selection of the processes and are not intended for comparison with existing market values.

3.6.4 **Product Costs** (cont'd)

The first, product cost on an "Energy Value" basis, is simply the unit energy cost ($\$/10^6$kJ) and is obtained by dividing the total annual costs by the total energy produced. The latter approach, "Product Value" method, was proposed and used by The Engineering Society for Commission on Energy (ESCOE)*. The objective of their method is to allow comparison to be made both on product quality as well as heating value. For example, gasoline has a higher market value than fuel oil and calculating the product cost solely on heating value would invert the price ranking of these fuels since heating oil has a higher fuel value per litre than does gasoline.

In the ESCOE methodology, a value factor F_i is defined as the ratio of market value of individual product M_i to the chosen reference product M_r, thus:

$$F_i = \frac{M_i}{M_r} = \frac{P_i}{P_r} \qquad (1)$$

Where, M_i = market value of individual product.
M_r = market value of the reference product.
P_i = cost of individual products, $\$/10^6$ kJ
and P_r = cost of reference product, $\$/10^6$ kJ

On an annual basis, a simplified cost equation may be represented as:

$$\sum P_i B_i = F + V + kC \qquad (2)$$

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* K.A. Rogers and R.F. Hill, "Coal Conversion Comparison", DOE Report No. FE-2468-51, July 1979

3.6.4 **Product Costs** (cont'd)

where, P_i = cost of individual product, \$/10^6kJ

B_i = annual production (energy output) of product "i" in 10^6kJ

F = annual fixed cost

V = annual variable cost

k = levelised capital cost factor to allow for interest, return and recovery of capital investment

C = total capital investment

Substitution the value of P_i from Equation (1) into Equation (2) gives:

$$P_r = \frac{F + V + kC}{\sum F_i B_i} \qquad (3)$$

Equation (3) is the basic cost equation relating the cost of a reference product P_r as a function of annual costs, annual production and the relative market values of various products. Thus, when P_r is determined, the cost of individual products on a "Value Basis" is obtained from Equation (1).

The product cost on an "Energy Basis", P_e is calculated by assigning the value of F_i as 1.0 for all the products in Equation (3) to obtain:

$$P_e = \frac{F + V + kC}{\sum B_i} \qquad (4)$$

where, $\sum B_i$ = total annual production in 10^6 kJ.

To calculate product cost P_r, a set of market values M_i is required to calculate value factors F_i of LSE products relative to premium gasoline M_r. The market values supplied by Petro-Canada is presented in Table 3-12. The market value are wholesale price and includes Federal Excise Tax. Provincial sales tax is not included.

Table 3-13 presents the production rate and annual energy output for LSE plant for both 65% and 82% extraction efficiencies. Using the operating, maintenance and production cost data given in Table 3-10, the product costs on "Energy Value" and "Product Value" basis are calculated and presented in Table 3-14.

Table 3-12 Value Factor of Products from LSE Plant

[Reference to Premium Gasoline in Saskatchewan]

Products	Market Value *	Converted Market Value**, $\$/10^6$ kJ	Value Factor, F_1
Propane/Butane (LPG)	\$19/bbl	5.01	0.42
Jet Fuel	\$35.62/bbl	7.33	0.61
Diesel Fuel	\$0.31/L	9.91	0.82
Unleaded Gasoline	\$0.344/L	11.54	0.96
Premium Gasoline (Reference Product)	\$0.359/L	12.04	1.00

* Market value incudes Federal Excise Tax but Provincial Sales Tax is excluded. Price for LPG and jet fuel are estimated values. Price for other products are wholesale price for Saskatoon, Saskatchewan.

** Market value supplied by PETRO-CANADA have been converted to a consistent unit basis ($\$/10^6$kJ) using conversion data given by ESCOE, DOE Report FE-2468-51, July 1929.

Source:

M. Taherzadeh, Coal and Alternate Fuel, PETRO-CANADA, Telex dated 7 April 1982 to J.A. Mikhlin, SNC Inc.

Table 3-13 Production Rate and Annual Energy Output for a LSE Plant

[Capacity: 25,000 t/d of Willow Bunch Lignite]
[310 operating days per year]

Extraction Efficiency	Products	Flow Rate		Heating Value	Energy Output, 10^{12} kJ/year	
		bbl/t	t/d	kJ/kg	$\sum B_i$	$\sum F_i B_i$
65% MAF Coal	Propane/Butane	4,157	389	45,100	5.43	2.28
	Jet Fuel	9,365	1,402	41,045	17.82	10.87
	Diesel Fuel	8,102	1,262	41,200	16.12	13.22
	Gasoline	10,296	1,490	40,120	18.54	17.80
					57.91	44.17
82% MAF Coal	Propane/Butane	4,382	410	45,100	5.74	2.41
	Jet Fuel	9,893	1,481	41,045	18.85	11.50
	Diesel Fuel	8,564	1,334	41,200	17.05	13.98
	Gasoline	10,863	1,572	40,120	19.53	18.75
					61.17	46.64

Table 3-14 Product Costs from LSE Plant

[Capacity: 25,000 t/d of Willow Bunch Lignite]
[mid-1982 Canadian dollars]

Reactor Extraction Efficiency		65% MAF Coal	82% MAF Coal
Average Product Cost, $\$10^{6}$kJ			
Energy Value Basis,		9.48	8.97
Product Value Basis,		12.43	11.76
Individual Product Costs (Product Value Basis), $/bbl			
Propane		22.03	20.84
Jet Fuel		46.59	44.08
Diesel Fuel		65.41	61.88
Unleaded Gasoline		69.28	65.55
Individual Product Costs (Product Value Basis)			
Propane	$/bbl	22.03	20.84
Jet Fuel	¢/L	29.30	27.70
Diesel Fuel	¢/L	41.10	38.90
Unleaded Gasoline	¢/L	43.60	41.20

* Reference to premium gasoline in Saskatchewan. Individual product cost in $\$/10^{6}$kJ relative to reference product may be obtained by multiplying the "Product Value" cost P_r by individual value factors F_i given in Table 3-12.

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REFERENCES AND SOURCES

REFERENCES AND SOURCES

General

H.I. Smith, "Coal Conversion Technology", Chemical Technology Review No. 66, Noyes Data Corporation, N.J., 1976.

J.D. Batchelor, U.S. Department of Energy, "Technical Status of the Large Coal Liquefaction Programs at the Department of Energy", paper presented at Institute of Gas Technology sponsored Symposium on Advances of Coal Utilization Technology, Louisville, Kentucky, May 14-18, 1979.

N. Berkowitz, "Introduction to Coal Technology", Academic Press, N.Y., 1979.

P. Nowacki, "Coal Liquefaction Processes", Chemical Technology Review No. 131, Energy Technology Review No. 45, published by Noyes Data Corp. N.J., 1979.

P. Sterbenz, "Coal Conversion Today, Hydrocarbon Processing, p.68-72, July 1980.

D.J. Deutsch, "Solid/Liquid Separation Routes are Nearing Demonstration Stage", Chemical Engineering, p.33-35, August 25, 1980.

B.R. Rodger et al., Oakridge National Laboratory, "Assessment of Current Research and Development in Support of the U.S. Coal Liquefaction Demonstration Plants Program", paper presented at the 15th Intersociety Energy Conversion Engineering Conference, Seattle, Washington, August 18-22, 1980, Proceedings Vol. 3, p.1812-1817.

L.M. Joseph, U.S. Department of Energy, "The Department of Energys' Major Project Coal Liquefaction Program", paper presented at the 15th Intersociety Energy Conversion Engineering Conference, Seattle, Washington, August 18-22, 1980. Proceedings Volume 3, p. 1808-1811.

H-Coal Process

E.S. Johanson, Hydrocarbon Research Inc., "Gas-Liquid Contacting Process", U.S. Patent No. 2,987,465 dated June 20, 1958 and U.S. Patent No. Re 25,770 dated April 27, 1965.

REFERENCES AND SOURCES (cont'd)

A.L. Conn et al., American Oil Company, Research and Development Department, "Evaluation of Project H-Coal", prepared for Department of Interior, Office of Coal Research, under Contract No. 14-01-0001-1188, December 1, 1967.

C.A. Johnson, M.C. Chervenak and E.S. Johanson, "Scaling-up Factors in the H-Coal Process", Chemical Engineering Progress Technical Manual on Coal Processing Technology, Vol 1, published by American Institute of Chemical Engineers, pp. 22-24, 1974.

H.H. Stotler, Hydrocarbon Research Inc., "The H-Coal Pilot Plant Program", Chemical Engineering Progress Technical Manual on Coal Processing Technology, Vol. 2, published by Americn Institute of Chemical Engineers, 1975.

Hydrocarbon Research Inc., "Horizontal Centrifuge Operation for H-Coal Liquid-Solid Separation", The United States Energy Research and Development Administration, Contract No. E(49-19) 0-1544, HRI Laboratory Report No. L-12-CL(350)-506, October 22, 1975.

B.C. Peters, The Dow Chemical Company, "Chemicals from Coal", prepared for U.S. Energy Research and Development Administration under Contract No. E(49-18)-1534, Report No. FE-1534-47, May 1977.

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant, Vol.1",Final Report of Phase I Study (December 1973-June 1976) prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1544, HRI Report No. HCP/T-1544/1, November 1977.

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant, Vol.1", Final Report of Phase I Study (December 1973-June 1976) prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1544, HRI Report No. HCP/T-1544/1, November 1977.

Hydrocarbon Research Inc., "H-Coal Intregrated Pilot Plant Vol.2", Final Report of Phase I Study (December 1973-June 1976) prepared for U.S. Department of Energy under Contract No.EX-76-C-01-1544, HRI Report No. HCP/T-1544/1, November 1977.

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant", Work Sponsored by Electric Power Research Institute under Contract EX-76-C-01-1544, EPRI Report No. AF-681, Project No. 238-1, March 1978.

REFERENCES AND SOURCES (cont'd)

A.L. Conn et al., American Oil Company, Research and Development Department, "Evaluation of Project H-Coal", prepared for Department of Interior, Office of Coal Research, under Contract No. 14-01-0001-1188, December 1, 1967.

C.A. Johnson, M.C. Chervenak and E.S. Johanson, "Scaling-up Factors in the H-Coal Process", Chemical Engineering Progress Technical Manual on Coal Processing Technology, Vol 1, published by American Institute of Chemical Engineers, pp. 22-24, 1974.

H.H. Stotler, Hydrocarbon Research Inc., "The H-Coal Pilot Plant Program", Chemical Engineering Progress Technical Manual on Coal Processing Technology, Vol. 2, published by Americn Institute of Chemical Engineers, 1975.

Hydrocarbon Research Inc., "Horizontal Centrifuge Operation for H-Coal Liquid-Solid Separation", The United States Energy Research and Development Administration, Contract No. E(49-19) 0-1544, HRI Laboratory Report No. L-12-CL(350)-506, October 22, 1975.

B.C. Peters, The Dow Chemical Company, "Chemicals from Coal", prepared for U.S. Energy Research and Development Administration under Contract No. E(49-18)-1534, Report No. FE-1534-47, May 1977.

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant, Vol.1",Final Report of Phase I Study (December 1973-June 1976) prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1544, HRI Report No. HCP/T-1544/1, November 1977.

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant, Vol.1", Final Report of Phase I Study (December 1973-June 1976) prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1544. HRI Report No. HCP/T-1544/1, November 1977.

Hydrocarbon Research Inc., "H-Coal Intregrated Pilot Plant Vol.2", Final Report of Phase I Study (December 1973-June 1976) prepared for U.S. Department of Energy under Contract No.EX-76-C-01-1544, HRI Report No. HCP/T-1544/1, November 1977.

Hydrocarbon Research Inc., "H-Coal Integrated Pilot Plant", Work Sponsored by Electric Power Research Institute under Contract EX-76-C-01-1544, EPRI Report No. AF-681, Project No. 238-1, March 1978.

REFERENCES AND SOURCES (cont'd)

T.O. Mitchell, Mobil Research and Development Corporation, "Response of Product and Solvent Composition to Catalyst during Coal Liquefaction", paper presented at the Fourth Annual Electric Power Research Institute Contractor's Conference on Coal Liquefaction, Palo Alto, California, May 8-10, 1978.

D.K. Jim and K.K. Robinson, Amoco Oil Research and Development, "Catalyst Development for Liquefying Eastern and Western Coal", paper presented at the Fourth Annual Electric Power Research Institute Contractors' Conference on Coal Liquefaction, Palo Alto, California, May 8-10, 1978.

A.G. Comolli, D.T.A. Huibers, E.S. Johanson and C.P. Laird, Hydrocarbon Research Inc., "Evaluation of Selective H-Coal Catalysts", paper presented at the fourth Annual Electric Power Research Institute Contractor's Conference on Coal Liquefaction", Palo Alto, California, May 8-10, 1978.

W. Karolkiewicz, Hydrocarbon Research Inc., "Solids-Liquid Separation in H-Coal Development", paper presented at AIChE meeting, Philadelphia, Pennsylvania, June 1978.

A.G. Commolli and E.S. Johanson, Hydrocarbon Research Institute "Bench-Scale Catalyst Evaluation Program", prepared for U.S. Department of Energy, U.S.A., under contract EX-77-C-01-2547, Report No. FE-12547-28, December 1978.

A.G. Comolli, E.S. Johanson and P. Sheth, Hydrocarbon Research Inc., "H-Coal PDU Tests on Illinois No.6 and Wyoming Coals", Special Report prepared for Electric Power Research Institute, EPRI Report No. AF-1143-SR, July 1979.

Hydrocarbon Research Inc., "H-Coal Run Summary of Bench-Scale and Process Development Unit Operations", Issued by HRI, R & D Centre, N.J., October 14, 1979.

Ashland Synthetic Fuels Inc., "H-Coal - A Future Energy Source", Information Sheet published by Ashland Oil Inc., November 1979.

Ashland Synthetic Fuels, Inc., "H-Coal Pilot Plant, Cattlettsburg, Fact Sheet", Information Sheet published by Ashland Oil Inc., November 1979.

P.A. Buckingham, R.Ravikumar, and A.A. Leavitt, Fluor Engineers and Constructor Inc., "Engineering Evaluation of Conceptual Coal Conversion Plant Using the H-Coal Liquefaction Process", prepared for Electric Power Research Institute under Contract No. RP-411-4, Report No. AF-1297, December 1979.

REFERENCES AND SOURCES (cont'd)

H.H. Stotler and R.T. Schutter, "Status and Plans of H-Coal Pilot Plant", Chemical Engineering Progress Technical Manual on Coal Procesing Technology, Vol.5, published by American Institute of Chemical Engineers, 1979.

The Energy Daily, National Press Building, Washington, Thursday, February 7, 1980.

J.A. Mikhlin, SNC Inc., Visit to Hydrocarbon Research Inc., H-Coal Experimental Facilities and Laboratories,Lawrenceville, N.J., March 1980.

Hydrocarbon Research Inc., "H-Coal Process", paper presented at Current Coal Conversion Technology Course at Center for Professional Advancement, East Brunswick, N.J., March 11, 1980.

The Energy Daily, National Press Building, Washington, Thursday, July 31, 1980.

J.A. Mikhlin, SNC Inc., Visit to H-Coal Pilot Plant at Catlettsburg, Kentucky, and meeting at HRI Inc., Lawrenceville, N.J., August 11-14, 1980.

Energy Review, Issued by U.S. Department of Energy, Energy Information Administration, September 1980.

Synfuels, McGraw Hill, N.Y., p. 4, September 5, 1980.

Synfuels, McGraw Hill, N.Y., pp. 1-2, October 3, 1980.

The Energy Daily, National Press Building, Washington, Tuesday, October 14, 1980.

Synfuels, McGraw Hill, N.Y., p. 8, October 14, 1980.

R.M.Eccles and G.R.DeVaux, Hydrocarbon Research Inc., "Current Status of H-Coal Commercialization", paper presented at the Syncrude Fuel Conference sponsored by Electric Power Research Institute and Ministry of Energy of the Federal Republic of Germany, San Francisco, California, October 13-16, 1980.

J.L. Gendler, HRI, Personal Communication to Dr. J.A. Mikhlin, SNC Inc., HRI Ref. No.01-JLG-17-81, January 23, 1981.

HRI Report, "H-Coal Evaluation for the Energy Authority of New South Wales", August 1981.

REFERENCES AND SOURCES (cont'd)

R.M. Eccles and G.R. DeVaux, Hydrocarbon Research Inc., "Current Status and Commercialization", paper presented at the 2nd World Congress of Chemical Engineering of Chemical Engineering held in Montreal, Canada, October 4-9, 1981. Proceeding Vol. 2, pp. 514-520.

D.J. O'Rear, R.F. Sullivan, and B.E. Stangeland, Chevron Research Company, "Catalytic Upgrading of H-Coal Syncrudes", published in "Upgrading Coal Liquids", edited by R.F. Sullivan, American Chemical Society Symposium Series No. 156, pp. 115-144, 1981.

Synfuels, McGraw Hill Inc., "H-Coal Plants' Liquefaction Section will be 35% of Total Project Cost", p. 4, November 1981.

The Energy Daily, National Press Building, Washington, "H-Coal Lays Plans for When the DOE Money Runs Out", January 19, 1982.

RAG/Veba Process

R. Specks and J. Langhoff, Ruhrkohle AG, and B.Cornils, Ruhrchemie AG, "The Development of a Pulverized Coal Gasifier in the Prototype - Phase Using the Texaco-System", paper presented at the Symposium on Coal Refining, Edmonton, Alberta, April 20-21, 1978.

H.E. Wurfel, Saarbergwerke AG, "I.G. Farben Process Further Development of a Technically Proven Process", paper presented at the Symposium on Coal Refining, Edmonton, Alberta, April 20-21, 1978.

B. Strobel, I. Romey and G. Kolling, Bergbau-Forschung GmbH, "German Routes to Coal Hydrogenation", paper presented at Institute of Gas Technology sponsored Symposium on Advances in Coal Utilization Technology at Louisville, Kentucky, May 14-18, 1979.

J.B. O'Hara, N.E. Jentz and A. Bela, The Ralph M. Parsons Company, "Coal Liquefaction - Impact of Product Composition on Relative Economics", paper presented at the Annual Fall Conference of the Canadian Society for Chemical Engineering at Sarnia, Ontario, 30 September - 3 October, 1979.

J. Langhoff, E. Wolowski, G.Escher, and H. Hosang, "Status of German Technology on Coal Hydrogenation Development by Ruhrkohle AG and Veba Oel AG", paper presented at the Annual Fall Conference of the Canadian Society for Chemical Engineering at Sarnia, Ontario, 30 September - 3 October, 1979.

REFERENCES AND SOURCES (cont'd)

Birtley Coal and Mineral Testing, and Cascade Coal Petrography Ltd., "Proximate Ultimate and Petrographic Analaysis of Canadian Subbituminous Coal from Forestburg, Alberta", analysis done for SNC Inc., Montreal, October 1979.

R. Duecker, Ruhrkohle AG, "Governmental and Industrial Measures in the Fields of Coal Gasification and Coal Liquefaction in the Federal Republic of Germany", paper presented at the Seventh Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity, held at University of Pittsburgh, Pennsylvania, August 5-8, 1980.

R. Specks, Ruhrkohle AG, "Gasification and Hydrogenation of Coal-Status and Prospects", paper presented at the Fifth International Conference on Coal Research held at Dusseldorf, Federal Republic of Germany, September 1-5, 1980.

G. Escher et al, "Ruhrkohle/Veba Oel Technology for Hydrogenation of Hard Coal", paper presented at Annual Fall Conference of the Canadian Society for Chemical Engineering, Edmonton, Alberta, October 19-22, 1980.

Synfuels, "Saarberg Coal Liquid Pilot Plant to Test New Heat Transfer Method", November 14, 1980, McGraw Hill Inc., New York.

B. Strobel and F. Friedrich, Bergbau-Forschung GmbH, "Liquefacton of Coal in the Federal Republic of Germany", paper presented at the Symposium of the Royal Academy of Engineering Sciences on Coal Liquefaction at Stockholm, December 2, 1980.

D. Oliver, Financial Times Business Information Ltd., "Oil and Gas from Coal", Financial Times Business Information Ltd., Bracken House, London, January 1981.

J.A. Mikhlin, SNC Inc., Report of the Ruhrkohle AG/Veba Oel AG/ SNC meeting on February 16, 1981, Essen, Federal Republic of Germany.

J.A. Mikhlin, SNC Inc., Visit to Beregbau-Forschung Coal Liquefaction Facilities, Essen, and 200 tonnes-per-day pilot plant at Bottrop, Federal Republic of Germany,February 16-17, 1981.

Veba Oel AG, Buer,Federal Republic of Germany, "Description and Flowsheet of the Bottrop Pilot Plant", February 20, 1981.

Veba Oel AG, Buer, Federal Republic of Germany, Technical Information on "Conceptual Design Data of Commercial Plant for Coal Hydrogenation", February 20, 1981.

REFERENCES AND SOURCES (cont'd)

Veba Oel AG, Buer, Federal Republic of Germany, Technical Information on "Upgrading of Oil from Coal", February 20, 1981.

Veba Oel AG, Buer, Federal Republic of Germany, Information Package and Report on "Autoclave Tests of the Liquefaction Behaviour of Canadian Subbituminous Coal", July 29, 1981.

J.A. Miklhin, SNC Inc., Visit to Veba Oel AG, October 29 - November 2, 1981.

Synfuels, McGraw Hill Inc., "West Germany Designing Incentives for Coal Liquefaction Project", pp. 3-4, January 29, 1982; and "West Germany Charts Out $215-million Synfuels Pilot Program", pp. 2-3, January 8, 1982.

Rheinbraun Process

R. Duecker, Ruhrkohle AG, "Government and Industrial Measures in the Fields of Coal Gasification and Coal Liquefaction in the Federal Republic of Germany", paper presented at the Seventh Annual International Conference on Coal Gasification, Liquefaction and Converesion to Electricity, held at University of Pittsburgh, Pennsylvania, August 5-8, 1980.

D. Boecker et al, Rheinische Braunkohlenwerke AG and Union Rheinische Braunkohlen Kraftstoff AG, "Rheinbraun Liquefaction of Brown Coal", paper presented et EPRI/KFA - Juklich Conference on Synthetic - Fuels: Status and Direction, at San Francisco, October 1980.

Synfuels, "Rheinbraun and Victoria Sign Agreement on Australian Coal Liquids Plant", December 12, 1980, McGraw Hill Inc., New York.

J.A. Mikhlin, SNC Inc., Visit to Rheinische Braunkohlenwerke AG, Koln, February 19, 1981; and October 29 - November 2, 1981.

D. Boecker, W. Dolkemeyer, K.-H. Keim, U. Lenz and G. Ritter, "Results of Liquefaction of Rhenish Brown Coal", paper presented at the "Lignite Symposium", held at San Antonio, June 1981.

EDS Process

ER and E Co, "EDS Coal Liquefaction Process Development, Phase IIIA, Final Technical Progress Report", January 1, 1976 - June 30, 1977, Vol. I and II. prepared for U.S. Energy Research and

REFERENCES AND SOURCES (cont'd)

Development Administration under the contract No. E(49-18) 2353, Report No. FE-2353-20, February 1978.

ER and E Co., "EDS Coal Liquefaction Process Development, Phase IV, Summary of EDS Predevelopment (1966-1975)", prepared for U.S. Department of Energy under the contract No. EF-77-A-01-2893, Report No. FE-2893-16, July 1978.

ER and E Co., "EDS Coal Liquefaction Process Development, Phase IV, Quarterly Technical Progress Report, July 1-Sept.30, 1978", prepared for U.S. Department of Energy under the Contract No. EF-77-A-01-2893, Report No. FE-2893-21, December 1978.

J.B. Stone, K.L. Trachte, and S.K. Poddar, "Calcium Carbonate Deposit Formation During the Liquefaction of Low Rank Coals", Proceedings of ACS Division of Fuel Chemistry, Vol.24, No.2, ACS National Meeting, Honolulu, Hawaii, April 2-6, 1979.

K.L. Trachte, "Liquefaction of Wyoming Subbituminous Coal with the EDS Process", paper presented at the 4th Annual Contractor's Conference of Electric Power Research Institute on Coal Liquefaction, Palo Alto, California, May 9-10, 1979.

S. Zaczepinski et al., Upgrading of Coal Liquids, paper presented at the mid-year API Meeting, San Francisco, California, May 14-17, 1979.

ER and E Co., "EDS Coal Liquefaction Process Development, Phase IV, Quarterly Technical Progress Report, January 1-March 31, 1979", prepared for U.S. Department of Energy under the Contract No. EF-77-A-01-2893, Report No. FE-2893-29, June 1979.

ER and E Co., "EDS Commercial Plant Study Design Update Offsite Design Basis - Illinois Coal Market Flexibility Sensitivity Case", prepared for U.S. Department of Energy under the contract No. EF-77-A-01-2893, Report No. FE-2893-37, July 1979.

ER and E Co., "EDS Coal Liquefaction Process Developoment, Phase IV, Annual Technical Progress Report, July 1, 1978 - June 30, 1979" prepared for U.S. Department of Energy under contract No. EF-77-A-01-2893, Report No. FE-2893-35, September 1979.

W.R. Epperly and J.W. Taunton, "Exxon Donor Solvent Progress Development" paper presented at 29th Canadian Chemical Engineering Conference, Sarnia, Ontario, September 30 - October 3, 1979.

REFERENCES AND SOURCES (cont'd)

D.F. Ryon, B.H.White, and S. Zaczepinski, Exxon research and Engineering Co. "Characterization and Processibility of EDS Product Streams, Proceedings of 5th EPRI Contractor's Conference on Coal Liquefaction, Palo Alto,CA, May 1980.

L.L. Ansell, J.W. Taunton and K.L.Trachte, "Bottoms Recycle Studies in the EDS Process Development", Proceedings of the 5th Electrical Power Research Institute Contractor's Conference on Coal Liquefaction, Palo Alto,California, May 1980.

W.R.Epperly, K.W. Plumlee, and D.T. Wade, Exxon Research and Engineering Company, Florham Park, N.J. "Exxon Donor Solvent Coal Liquefaction Process; Development Program Status", Proceedings of 15th Intersociety Energy Convesion Engineering Conference, Seattle, Washington, August 18-22, 1980.

A.S.Crocker "EDS Commercial Plant Study Design Update Comparison of Base Coal with Market Sensibility Case Economics Sensitivities", Synthetic Fuel Research,Exxon Research and Engineering Company, Florham Park, N.J. August 27, 1980.

J.A. Mikhlin, Visit to Exxon Research and Engineering Company and EDS Experimental Facilities at Florham Park, N.J. and Pilot Plants, Baytown, Texas, August 27-28, 1980.

W.R. Epperly, K.W. Plumlee, and D.T. Wade, "Exxon Donor Solvent Coal Liquefaction Process, Development Program Status II", presented at the 5th Annual Conference on Materials for Coal Conversion and Utilization, Nationals Bureau of Standards, Caithersburg, Maryland, October 7-9, 1980.

ER and E Co. "EDS Coal Liquefaction Process Delvelopment Phase V, Quarterly Technical Progress Report, July 1 - September 30, 1980", prepared by U.S. Department of Energy under the contract No. DE-FC01-77ET10069, Report No. FE-2893-57, February 1981.

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal, Volume 1 - Main Report", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET 10069, Report No. FE-2893-61 (Volume 1), March 1981.

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal, Volume 2 - Base Case", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-61, Vol. 2, March 1981.

REFERENCES AND SOURCES (cont'd)

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal,Volume 3 - Market Flexibility Sensitivity Case", prepared for U.S. Department of Eknergy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-61, Vol.3, March 1981.

R.H.Brackett, T.J. Clunie, and A.M. Goldstein, Exxon Company, U.S.A. and Exxon Research and Engineering Company, Florham Park, N.J., "Exxon Donor Solvent Coal Liquefaction Process: Exxon Coal Liquefaction Plant Operations", presented at 6th Annual EPRI Contractor's Conference on Coal Liquefaction, May 13-14, 1981.

W.R. Epperly, K.W. Plumlee, and D.T. Wade, "Exxon Donor Solvent Coal Liquefaction Process: Development Program Status VI", Chemical Engineering Progress, May 1981.

The Energy Daily, National Press Building, Washington,D.C., Friday, June 19, 1981.

S.J. Hsia, K.L.Trachte, G.A. Wilcox and R.D. Williams, "Liquefaction of Low Rank Coals with the Exxon Donor Solvent Coal Liquefaction Process", presented at the 11th Biennial Lignite Symposium, San Antonio, Texas, June 15-17, 1981.

ER and E Co., "EDS Coal Liquefaction Process Development, Phase V, Quarterly Technical Progress Report, January 1 - March 31, 1981", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-70, Vol.1, July 1981.

J.W. Taunton, K.L. Trachte, and R.D. Williams, "Coal Feed Flexibility in the Exxon Donor Solvent Coal Liquefaction Process", FUEL, vol.60, September 1981.

A.M. Goldstein, "The Exxon Donor Solvent Process: Equipment and Scale-up Issues", paper presented at the Second World Congress of Chemical Engineering, Montreal, Canada, October 4-9,1981.

J.W. Taunton, K.L. Trachte, and R.D. Williams, Exxon Research and Engineering Company, "Coal Feed Flexibility in the Exxon Donor Solvent Coal Liquefaction Process" 1981.

ER and E Co., "Analysis of Alternative Processes for Manufacture of Fuel and Hydrogen in a Coal Liquefaction Plant - The EDS Process", Draft Report, 1981.

REFERENCES AND SOURCES (cont'd)

D.F. Ryon, B.H.White, and S. Zaczepinski, Exxon research and Engineering Co. "Characterization and Processibility of EDS Product Streams, Proceedings of 5th EPRI Contractor's Conference on Coal Liquefaction, Palo Alto,CA, May 1980.

L.L. Ansell, J.W. Taunton and K.L.Trachte, "Bottoms Recycle Studies in the EDS Process Development", Proceedings of the 5th Electrical Power Research Institute Contractor's Conference on Coal Liquefaction, Palo Alto,California, May 1980.

W.R.Epperly, K.W. Plumlee, and D.T. Wade, Exxon Research and Engineering Company, Florham Park, N.J. "Exxon Donor Solvent Coal Liquefaction Process; Development Program Status", Proceedings of 15th Intersociety Energy Convesion Engineering Conference, Seattle, Washington, August 18-22, 1980.

A.S.Crocker "EDS Commercial Plant Study Design Update Comparison of Base Coal with Market Sensibility Case Economics Sensitivities", Synthetic Fuel Research,Exxon Research and Engineering Company, Florham Park, N.J. August 27, 1980.

J.A. Mikhlin, Visit to Exxon Research and Engineering Company and EDS Experimental Facilities at Florham Park, N.J. and Pilot Plants, Baytown, Texas, August 27-28, 1980.

W.R. Epperly, K.W. Plumlee, and D.T. Wade, "Exxon Donor Solvent Coal Liquefaction Process, Development Program Status II", presented at the 5th Annual Conference on Materials for Coal Conversion and Utilization, Nationals Bureau of Standards, Caithersburg, Maryland, October 7-9, 1980.

ER and E Co. "EDS Coal Liquefaction Process Delvelopment Phase V, Quarterly Technical Progress Report, July 1 - September 30, 1980", prepared by U.S. Department of Energy under the contract No. DE-FC01-77ET10069, Report No. FE-2893-57, February 1981.

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal, Volume 1 - Main Report", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET 10069, Report No. FE-2893-61 (Volume 1), March 1981.

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal, Volume 2 - Base Case", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-61, Vol. 2, March 1981.

REFERENCES AND SOURCES (cont'd)

ER and E Co., "EDS Commercial Plant Study Design Update - Illinois Coal,Volume 3 - Market Flexibility Sensitivity Case", prepared for U.S. Department of Eknergy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-61, Vol.3, March 1981.

R.H.Brackett, T.J. Clunie, and A.M. Goldstein, Exxon Company, U.S.A. and Exxon Research and Engineering Company, Florham Park, N.J., "Exxon Donor Solvent Coal Liquefaction Process: Exxon Coal Liquefaction Plant Operations", presented at 6th Annual EPRI Contractor's Conference on Coal Liquefaction, May 13-14, 1981.

W.R. Epperly, K.W. Plumlee, and D.T. Wade, "Exxon Donor Solvent Coal Liquefaction Process: Development Program Status VI", Chemical Engineering Progress, May 1981.

The Energy Daily, National Press Building, Washington,D.C., Friday, June 19, 1981.

S.J. Hsia, K.L.Trachte, G.A. Wilcox and R.D. Williams, "Liquefaction of Low Rank Coals with the Exxon Donor Solvent Coal Liquefaction Process", presented at the 11th Biennial Lignite Symposium, San Antonio, Texas, June 15-17, 1981.

ER and E Co., "EDS Coal Liquefaction Process Development, Phase V, Quarterly Technical Progress Report, January 1 - March 31, 1981", prepared for U.S. Department of Energy under cooperative agreement No. DE-FC01-77ET-10069, Report No. FE-2893-70, Vol.1, July 1981.

J.W. Taunton, K.L. Trachte, and R.D. Williams, "Coal Feed Flexibility in the Exxon Donor Solvent Coal Liquefaction Process", FUEL, vol.60, September 1981.

A.M. Goldstein, "The Exxon Donor Solvent Process: Equipment and Scale-up Issues", paper presented at the Second World Congress of Chemical Engineering, Montreal, Canada, October 4-9,1981.

J.W. Taunton, K.L. Trachte, and R.D. Williams, Exxon Research and Engineering Company, "Coal Feed Flexibility in the Exxon Donor Solvent Coal Liquefaction Process" 1981.

ER and E Co., "Analysis of Alternative Processes for Manufacture of Fuel and Hydrogen in a Coal Liquefaction Plant - The EDS Process", Draft Report, 1981.

REFERENCES AND SOURCES (cont'd)

LSE Process

G.A.M.Diepen and F.E.C.Scheffer, J.AM. Chem. Soc., Vol.70, 4085, 1948.

J.C. Whitehead and D.F.Williams, National Coal Board, J. Inst. Fuel, p. 182 (1971)

K.D. Bartle et al., Fuel, Vol.54, No.226 (1975.

R.R. Maddocks, Catalytic Inc., J.Gibson, Coal Utilisation and Research,"Supercritical Extraction of Coal", Chem. Eng. Progr., pp 59-63, June 1977.

M.E. Crowther and J.C. Whitehead, National Coal Board, "Fluidisation of Fine Particles at Elevated Pressure", Fluidisation, pp. 65-70, 1978.

"U.K.Expands R 8 D on Chemicals from Coal", Chemical and Engineering News, pp. 10-11, June 19, 1978.

G.O.Davies, W. Wyss and D.G. Gavin, "Making Substitute Distillates from Coal for the Petroleum and Chemical Industries", EEC Symposium on Gasification and Liquefaction of Coal, Katowice, Poland, April 1979.

M.D. Gray, Coal Processing Consultants Ltd., "Liquids from Coal Processes Under Development by the National Coal Board," UK, paper presented at 29th Canadian Chemical Engineering Conference on Synthetic Liquids from Coal and Biomass, held at Sarnia,Ontario, September 30 to October 3, 1979.

J.A. Mikhlin, SNC Inc., Visit to Experimental Facilities of Coal Research Establishment, National Coal Board, February 1981.

G.G. Thurlow, National Coal Board, "Producing Fuels and Chemical Feedstocks from Coal", Chem. Eng. Progr. pp. 81-84, March 1980.

J.C. Clarke and T.D.Rantell, National Coal Board, "Filtration in Coal Liquefaction, Influence of Digestion Conditions in the Filtration of Non-Hydrogenated Coal Digests", Fuel, Vol.59, pp.208-212, March 1980.

Synfuels, McGraw Hill,N.Y., "U.K. Planning Massive Coal-Based Synfuels Program by Early Next Century", pp.1 and 8, August 22, 1980.

REFERENCES AND SOURCES (cont'd)

Energy Review, U.S. Department of Energy, Energy Information Administration, September 1980.

National Coal Board, "Coal Liquefaction for Transport Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", report prepared by NCB, Coal Research Establishment, Cheltenham, February 1981.

D.F. Williams, Head of Coal Refining, National Coal Board, Personal Communi[illegible] to Dr. J.A. Mikhlin, SNC Inc., dated February 25, 1981

National Coal Board, "Coal Liquefaction for Transport Fuels Using the Supercritical Gas Extraction Process of the U.K. National Coal Board", report prepared by NCB for SNC Inc., March 1981.

J. Owen,National Coal Board, "Conversion and Uses of Liquid Fuels from Coal", Fuel, Vol 60, pp. 755-761, September 1981.

P. Cammack, National Coal Board, "Coal Preparation for Coal Conversion Processes", Fuel, Vol 60, pp 762-767, September 1981.

W.P.Syrett, National Coal Board, Personal Communication to J.A. Mikhlin, SNC Inc., "The Processing of Saskatchewan Lignites by the NCB's Liquid Solvent Extraction Process", NCB Reference No. DR3000, DR496/2, October 1981.

Lurgi/Ruhrgas Process

R.W.Rammler "Practical Experience with the Lurgi-Ruhrgas Process for the Devolatilization of Fine-Granular Fuel", Translation of an particle published in Erdol und Kohle, Erdgas, Petrochemie, 19, pp 117-121, 1966.

R.W. Rammler, "The Retorting of Coal, Oil Shale and Tar-Sand by Means of Circulated Fine-Grained Heat Carriers as a Preliminary Stage in the Production of Synthetic Crude Oil", paper presented at the AIME Annual Meeting, Hydrocarbon Session, Denver, Colorado, Feb. 15-19, 1970.

P.F.H. Rudolph and P.K. Hebert, Lurgi Kohle und Mineraloltechnik GmbH, "Conversion of Coal to High Value Products", Lurgi Express Information, Ref. No. 0152/9.75, 1975.

R.W. Rammler, "Retorting of Coal, Oil Shale and Tar Sand by Means of the Lurgi-Ruhrgas Process" modified version of the paper presented at the German-Canadian Coal Refining Symposium, Edmonton, Alberta, April 20-21, 1978.

REFERENCES AND SOURCES (cont'd)

H. Hiller, "Energy Perspectives", Lurgi Express Information, Ref. No. 01336/9.79, 1979.

P. Nowacki, "Coal Liquefaction Processing", Chemical Technology Review No. 131, Energy Technology Review No. 45, Noyes Data Corp, 1979.

R.W. Rammler, "The Lurgi-Ruhrgas Process for Carbonization of Coal", paper presented at the EPRI Conference on Coal Pyrolysis, Palo Alto, February 25-26, 1981.

D.N. Perkins, "Production of Liquid Hydrocarbons from Coal and its Impact on the Electricity Supply, Industry in Queensland", Proceedings of the EPRI Conference on Coal Pyrolysis, Palo Alto, February 25-26, 1981.

Z. Kolisnyk and R.A. Grant, "Coal Pyrolysis - A Third Alternative", paper presented at 33rd Canadian Chemical Engineering Conference on Coal", Regina, Saskatchewan, September 20-21, 1981.

B.C. Research Laboratories, "Liquefaction and Pyrolysis Tests of Samples of Estevan and Coronach Lignites, Project Report 2-02-691, 1982.

Techno-Economic Assessment of LSE Process

"Development Document for Effluent Guidelines and New Source Performance Standards for the Petroleum Refining Point Source Category", U.S. Environmental Protection Agency, EPA Report No. 440/1-74-014A, Washington D.C., April 1974.

Environment Saskatchewan, Air Pollution Control Regulations, 211/75, 1975.

Environment Saskatchewan, Water Pollution Control Branch, "Water Quality Objectives", January 1977, February 1977.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job No. C 8093, October 30, 1977.

B.C. McBeath, The Engineering Societies Commission on Energy Inc., "Guidelines for Economic Evaluation of Coal Conversion Processes, prepared for the U.S. Department of Energy under Contract No. EF-77-C-01-2468, Report No. FE-2468-44, April 1979.

REFERENCES AND SOURCES (cont'd)

K.A. Rogers and R.F. Hill, The Engineering Societies Commission on Energy Inc., "Coal Conversion Comparison, prepared for the U.S. Department of Energy Inc. under Contract No. EF-C-01-2468, Report No. FE-2468-51, July 1979.

M.D. Gray, Coal Processing Consultant Ltd., "Liquids from Coal Processes Under Development by the National Coal Board, U.K.", paper presented at 29th Canadian Chemical Engineering Conference on Synthetic Liquids from Coal and Biomass, held at Sarnia, Ontario, September 30 - October 3, 1979.

National Coal Board Report to SNC Inc., "Coal Liquefaction for Transportation Fuels Using the Solvent Extraction Technology of the U.K. National Coal Board", February 1981.

R.M. Wham et al, "Liquefaction Technology Assessment - Phase 1: Indirect Liquefaction of Coal to Methanol and Gasoline Using Available Technology", Oak Ridge National Laboratory Report ORNL-5664, February 1981.

K.A. Rogers, U.S. Synthetic Fuels Corp, "Synfuel Technologies and Costs" paper presented at the AIME Meeting, Chicago, Illinois, February 24, 1981, published in Material and Society, Vol.5, No.4, pp.355-363, 1981.

R. Salmon et al, Oak Ridge National Laboratory, "Technical and Economic Evaluation of Coal Liquefaction Technologies - Phase 0: Preliminary Screening Evaluation", Report No. ORNL/TM-7581, April 1981.

Construction Price Statistics, "Chemical and Mineral Process Plant Indexes", Table 19; and "Union Wage Rate Index", Table 3; November 1981, Catalogue No. 62-007, published by Statistics Canada, Ottawa.

F. Mourits, Coal Liquefaction Study Management, Saskoil, letter to J.A. Mikhlin, SNC Inc., December 17, 1981.

Engineering New Record, January 7, 1982, January 16, 1982.

W.P.Syrett, National Coal Board, report to SNC Inc. "The Processing of Saskatchewan Lignite by the NCB's Liquid Solvent Extraction Process: Willow Bunch Lignite", NCB Reference No., 496/2, 7/3, January 23, 1982. (The report is attached in confidential Volume V of the present study, "Process Developers Data and Information").

G. Kimber, Coal Research Establishment National Coal Board, telex to J.A. Mikhlin, SNC Inc., February 22, 1982.

REFERENCES AND SOURCES (cont'd)

M. Taherzadeh, Coal and Alternate Fuels, PETRO CANADA, Telex to J.A. Mikhlin, SNC Inc., 7 April 1982.

Construction Collective Agreement Summaries, "Labour Rate in Saskatchewan", published by the Canadian Construction Association, Ottawa, April 1982.

National Coal Board report to SNC Inc., "The Assessment of a Canadian Lignite for the National Coal Board Liquid Solvent Extraction Process", April 1982. (Report is attached in confidential Volume V of the present study "Process Developers Data and Information").

H. Staege, Krupps Koppers GmbH, "Preliminary Study for the Production of Pure Methanol or Synthesis Gas from Coal in the Basis of the Kopper-Totzek Coal Gasification Process", Report to SNC Inc., Krupp Koppers Ref. No. VK-1/Mn-Ko-1813, June 1982. (The report is attached in the confidential Volume V of the present study, "Process Developers Data and Information").

APPENDIX A

TYPICAL ULTIMATE, PROXIMATE AND PETROGRAPHIC ANALYSES OF SASKATCHEWAN LIGNITES

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Table A-1 Typical Ultimate, Proximate, and Petrographic Analysis of Saskatchewan Lignites

	Estevan	Willow Bunch
Moisture (as received), wt%	35.0	35.0
Gross Calorific Value (GCV) kJ/kg MF Coal	24,270	20,270
Proximate Analysis, wt% MF Coal		
Volatile Matter	42.0	39.6
Fixed Carbon	42.0	38.5
Ash	15.4	21.0
Sulfur	0.6	0.9
Ultimate Analysis, wt% MF Coal		
Carbon	61.2	52.1
Hydrogen	4.5	4.0
Nitrogen	1.0	0.5
Sulfur	0.6	0.9
Oxygen	17.3	21.5
Ash	15.4	21.0
Petrographic Analysis, vol%		
Vitrinite*	81.2	77.6
Exinite	7.9	1.7
Inertinite	9.9	18.7
Mineral Matter	0.9	2.0
Ash Analysis, wt%		
Silica, SiO_2	47.1	42.1
Alumina, Al_2O3	17.8	15.6
Titania, TiO_2	**	0.9
Ferric Oxide, Fe_2O_3	2.5	6.7
Lime, CaO	11.7	14.3
Magnesia, MgO	2.9	5.0
Potassium Oxide, K_2O	0.5	1.3
Sodium Oxide, Na_2O	5.0	1.8
Phosphorous Pentoxide, P_2O_5	1.0	
Sulfur Trioxide, SO_3	1.0	12.3
Undetermined	10.5	

* Equivalent to huminite in low-rank coals.

** Undetermined.

Sources: Information package received from Saskoil at the Project Kick-off Meeting, Sept. 10-11, 1981

Letter from Dr. F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to Dr. J.A. Mikhlin, SNC Inc.

APPENDIX B

ULTIMATE AND PROXIMATE ANALYSES OF DIFFERENT RANK COALS

Table B-1 Ultimate and Proximate Analyses of Different Rank Coals

Process Coal(s)	H-Coal		
	North Dakota Lignite[1]	Australian Brown Muswellbrook[2]	Subbituminous Wyodak[3]
Moisture (as received), wt%	35.0	30.0	30.0
Gross Calorific Value (GCV) kJ/kg MF Coal	25,335	23,260	26,563
Proximate Analysis, wt% MF coal			
Volatile Matter	44.0	30.6	44.6
Fixed Carbon	45.8	46.8	46.8
Ash	10.2	22.6	8.6
Ultimate Analysis, wt% MF coal			
Carbon	65.3	63.4	66.8
Hydrogen	4.5	3.7	5.0
Nitrogen	0.8	1.5	0.8
Sulfur	1.1	0.6	1.0
Oxygen	18.0	8.2	17.7
Ash	10.3	22.6	8.7
Petrographic Analysis, vol%			
Vitrinite		69.0	
Exinite	N/A	5.0	N/A
Inertinite		19.0	
Minerals Matter		2.0	
Ash Analysis, wt%			
Silica, SiO_2			29.3
Alumina, Al_2O3			18.8
Titania, TiO_2			1.1
Ferric Oxide, Fe_2O_3			6.6
Lime, CaO	N/A	N/A	18.8
Magnesia, MgO			4.5
Potassium Oxide, K_2O			0.4
Sodium Oxide, Na_2O			0.6
Phosphorous Pentoxide, P_2O_5			0.4
Sulfur Trioxide, SO_3			17.3
Undetermined			1.8

Notes/Sources:

1. Coal supplied by Peabody Coal Company from Dickinson Area, Stark County, North Dakota (HRI Report, Project H-Coal", 1967)
2. HRI Report, "H-Coal Evaluation for the Energy Authority of New South Wales", August 1981
3. EPRI Report No. AF-1143-SR, July 1979.

Table B-1 Ultimate and Proximate Analyses of Different Rank Coals (cont'd)

Process	H-Coal	RAG/Veba	
Coal(s)	Bituminous Illinois No. 6[1]	Bituminous German[2]	Subbituminous Forestburg[3]
Moisture (as received), wt%	11.0	4.3	25.6
Gross Calorific Value (GCV) kJ/kg MF Coal	29,354	33,010	28,130
Proximate Analysis, wt% MF coal			
Volatile Matter	37.9	36.2	41.2
Fixed Carbon	51.5	59.3	53.6
Ash	10.6	4.5	5.2
Ultimate Analysis, wt% MF coal			
Carbon	70.9	80.5	70.5
Hydrogen	4.9	4.7	4.4
Nitrogen	1.5	1.6	1.3
Sulfur	3.5	1.6	0.5
Oxygen	8.6	7.1	18.1
Ash	10.6	4.5	5.2
Petrographic Analysis, vol%			
Vitrinite	N/A	64.0	71.4
Exinite		17.0	7.3
Inertinite		14.0	18.2
Mineral Matter		5.0	3.1
Ash Analysis, wt%			
Silica, SiO_2	47.8	32.9	39.7
Alumina, Al_2O_3	18.1	27.6	14.6
Titania, TiO_2	0.9	0.8	8.5
Ferric Oxide, Fe_2O_3	17.8	32.2	30.3
Lime, CaO	7.4	2.3	4.5
Magnesia, MgO	0.9	1.0	1.0
Potassium Oxide, K_2O	1.8	1.0	0.9
Sodium Oxide, Na_2O	0.5	1.5	0.2
Phosphorous Pentoxide, P_2O_5	0.01	0.6	0.4
Sulfur Trioxide, SO_3	3.7	–	–
Undetermined	1.1	0.1	–

Notes/Sources:

1. From Burning Star Mine, Illinois (EPRI Report No. AF-1143-SR, July 1979)
2. From Ruhr area, F-R Germany; also referred to as "standard" German coal by the process developer. (Langhoff et al, paper presented at 29th Canadian Chemical Engineering Conference at Sarnia, Sept. 30, 1979)
3. Coal analysis performed by Birtley Coal and Mineral Testing, and Cascade Coal Petrography Ltd., for SNC Inc., Oct. 1979

Table B-1 Ultimate and Proximate Analyses of Different Rank Coals (cont'd)

Process	Rheinbraun	EDS	
Coal(s)	Rhenish Brown Coal[1]	N. Dakota Indian Head[2]	Texas Big Brown[2]
Moisture (as received), wt%	60.0	N/A	33.6
Gross Calorific Value (GCV) kJ/kg MF coal	23,260	N/A	22,100
Proximate Analysis, wt% MF coal			
Volatile Matter	N/A	44.1	44.4
Fixed Carbon	N/A	45.7	39.2
Ash	3.2	10.2	16.4
Ultimate Analysis, wt% MF coal			
Carbon	65.8	63.8	62.0
Hydrogen	5.0	4.7	4.8
Nitrogen	0.7	19.2	14.5
Sulfur	0.3	0.9	1.1
Oxygen	25.0	1.2	1.2
Ash	3.2	10.2	16.4
Petrographic Analysis, vol%			
Vitrinite		80.5	65.3
Exinite	N/A	2.7	16.3
Inertinite		16.8	18.4
Mineral Matter			
Ash Analysis, wt%			
Silica, SiO_2		26.3	40.6
Alumina, Al_2O3		10.7	14.4
Titania, TiO_2		0.5	1.2
Ferric Oxide, Fe_2O_3		14.0	7.8
Lime, CaO	N/A	16.0	13.2
Magnesia, MgO		4.9	3.7
Potassium Oxide, K_2O		0.9	0.7
Sodium Oxide, Na_2O		8.0	0.5
Phosphorous Pentoxide, P_2O_5		0.4	0.6
Sulfur Trioxide, SO_3		18.4	16.8
Undetermined		-	0.4

Notes/Sources:

1. J.A. Mikhlin, SNC Inc., visit to Rheinbraun Facilities at Koln, Feb. 19, 1981
2. DOE Report FE-2893-35, p. 18, Sept. 1979

Table B-1 Ultimate and Proximate Analyses of Different Rank Coals (cont'd)

Process	LSE		L-R
Coal(s)	U.K. Bituminous CRC 702[1]	Australian Brown[2]	German Subbituminous
Moisture (as received), wt%	10.0	45.0	4.9
Gross Calorific Value (GCV) kJ/kg MF coal	34,300	23,260	N/A
Proximate Analysis, wt% MF coal			
Volatile Matter	37.7	41.4	38.5
Fixed Carbon	51.2	41.4	56.6
Ash	11.1	17.2	4.9
Ultimate Analysis, wt% MF coal			
Carbon	74.7	62.1	N/A
Hydrogen	4.8	3.6	
Nitrogen	1.7	0.4	
Sulfur	0.6	0.2	
Oxygen	7.1	16.5	
Ash	11.1	17.2	
Petrographic Analysis, vol%			
Vitrinite	N/A	N/A	N/A
Exinite			
Inertinite			
Mineral Matter			
Ash Analysis, wt%			
Silica, SiO_2	N/A	N/A	N/A
Alumina, Al_2O3			
Titania, TiO_2			
Ferric Oxide, Fe_2O_3			
Lime, CaO			
Magnesia, MgO			
Potassium Oxide, K_2O			
Sodium Oxide, Na_2O			
Phosphorous Pentoxide, P_2O_5			
Sulfur Trioxide, SO_3			
Undetermined			

Notes/Sources:

1. NCB Report, "Coal Liquefaction for Transport Fuel Using the Solvent Extraction Technology of U.K. National Coal Board", Feb. 1981
2. W.P. Syrett, NCB, Personal Communication to J.A. Mikhlin, SNC Inc., NCB Ref. DR 3000, DR 496/2, Oct. 1981
3. Rammler, R.W., "The Lurgi-Ruhrgas Process for Carbonization of Coal", paper presented at the EPRI Conference on Coal Pyrolysis, Palo Alto, California, Feb.25-26, 1981.

APPENDIX C

TYPICAL ANALYSES AND PROPERTIES OF WILLOW BUNCH LIGNITE CONSIDERED FOR TECHNO-ECONOMIC ASSESSMENT OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS

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Appendix C-1 Proximate Analysis and Gross Calorific Value of Willow Bunch Lignite (Coronach Coal)

	Average Coal Analysis (206 Samples)		
	As-Received	MF	MAF
Proximate Analysis, wt%			
Moisture	35.0	-	-
Volatile Matter	25.7	39.6	50.1
Fixed Carbon	25.0	38.5	48.8
Ash	13.7	21.0	-
Sulfur	0.6	0.9	1.1
Gross Calorific Value, kJ/kg	13260	20380	25848

	Worst Coal Analysis		
	As-Received	MF	MAF
Proximate Analysis, wt%			
Moisture	40.0	-	-
Volatile Matter	20.4	34.0	50.0
Fixed Carbon	19.9	33.2	48.8
Ash	19.2	32.0	-
Sulfur	0.6	0.8	1.2
Gross Calorific Value, kJ/kg	10476	17461	25995

	Best Coal Analysis		
	As-Received	MF	MAF
Proximate Analysis, wt%			
Moisture	30.0	-	-
Volatile Matter & Fixed Carbon	59.5	85.0	99.2
Ash	10.0	14.3	-
Sulfur	0.5	0.7	0.8
Gross Calorific Value, kJ/kg	15817	22594	26581

Sources: Letter from Dr. F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to Dr. J.A. Mikhlin, SNC Inc.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job No. C-8093, October 30, 1977.

Appendix C-2 Ultimate Analysis of Willow Bunch Lignite (Coronach Coal)

	Average Coal Analysis (206 samples)	Worst Coal
Ultimate Analysis, wt% MF Coal		
Carbon	52.1	45.0
Hydrogen	4.0	3.5
Nitrogen	0.5	0.3
Sulfur	0.9	0.8
Oxygen	21.5	18.4
Ash	21.0	32.0

Sources: Letter from Dr. F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to Dr. J.A. Mikhlin, SNC Inc.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job No. C-8093, October 30, 1977.

Appendix C-3: Ash Analysis of Willow Bunch Lignite (Coronach Coal)

	In Oxidizing Atmosphere (108 Samples)	In Reducing Atmosphere (69 Samples)
Ash Fusion Temperature, °C		
Initial Deformation	1153	1173
Softening Temperature	1211	1192
Hemispherical Temperature	1236	1199
Fluid Temperature	1293	1209

	Average (133 Samples)
Chemical Analysis of the Ash	
Silica, SiO_2	42.1
Alumina, Al_2O_3	15.6
Ferric Oxide, Fe_2O_3	6.7
Lime, CaO	14.3
Magnesia, MgO	5.0
Potassium Oxide, K_2O	1.3
Sodium Oxide, Na_2O	1.8
Titania, TiO_2	0.9
Balance	12.3

Ash Distribution No. of Samples	wt% MF Coal
2	5 - 10
33	10 - 15
35	15 - 20
33	20 - 25
13	25 - 30
8	30 - 35
5	35 - 40
4	40

Sources: Letter from Dr. F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981, to Dr. J.A. Mikhlin, SNC Inc.

Saskatchewan Power Corporation, Research and Development Centre, "Coronach Coal Gasification Study", Volume II, Job No. C-8093, October 30, 1977.

snc

APPENDIX D

PRELIMINARY CHARACTERIZATION OF WILLOW BUNCH LIGNITE (CORONACH # 3216) FROM SASKATCHEWAN, CANADA, IN LIQUID SOLVENT EXTRACTION PROCESS

Appendix D-1 Results of Micro-autoclave Tests on Willow Bunch Lignite (Coronach #3216) Carried Out at Coal Research Establishment, National Coal Board.

ATT. DR. J.A. MIKHLIN
SECTION HEAD,
COAL R AND D,
SNC INC
MONTREAL.

WE HAVE EXTRACTED WILLOWBUNCH COAL AND EXTRACTION LEVEL HAS BEEN IN RANGE 60-70□ DMMF COAL. OUR PETROGRAPHIC EXAMINATION SUGGESTS THAT THIS IS DUE TO A HIGH LEVEL OF INERTINITE AND WE NOTE THAT VALUES SUPPLIED BY YOU ON ESTERAN DEPOSITS VARY BETWEEN 1 AND 45□, THIS IS AN EXCEPTIONALLY WIDE VARIATION BY UK LOW RANK COAL STANDARDS. IT WOULD SEEM THAT, AS WITH BITUMINOUS COALS, THE INERTINITE IS THE MOST DIFFICULT PART TO EXTRACT, SO MONITORING A LARGE DEPOSIT BY PETROGRAPHIC ANALYSES OR OUR EXTRACTION TEST IS ADVISED- THIS LEVEL OF EXTRACTION WILL APPLY TO SIMILAR DIRECT LIQUEFACTION PROCESSES, EVEN IF IT COULD BE EXTRACTED MUCH MORE HYDROGEN WILL BE NEEDED TO CONVERT THIS INERTINITE EXTRACT WITH LIGHT LIQUIDS.

HOWEVER DESPITE THE HIGHER FILTER CAKE YIELD THE LSE PROCESS CAN OPERATE WITH WILLOWBUNCH COAL WITH ONLY A 2□ DROP IN OVERALL THERMAL EFFICIENCY COMPARED WITH THE CASE SENT TO YOU EARLIER BY MR. SYRETT, THE FILTER CAKE JUST ABOUT SATISFIES THE REQUIREMENTS FOR GASIFICATION AND POWER GENERATION SO LITTLE ADDITIONAL COAL OTHER THAN THAT FED TO EXTRACTION WOULD BE NEEDED.

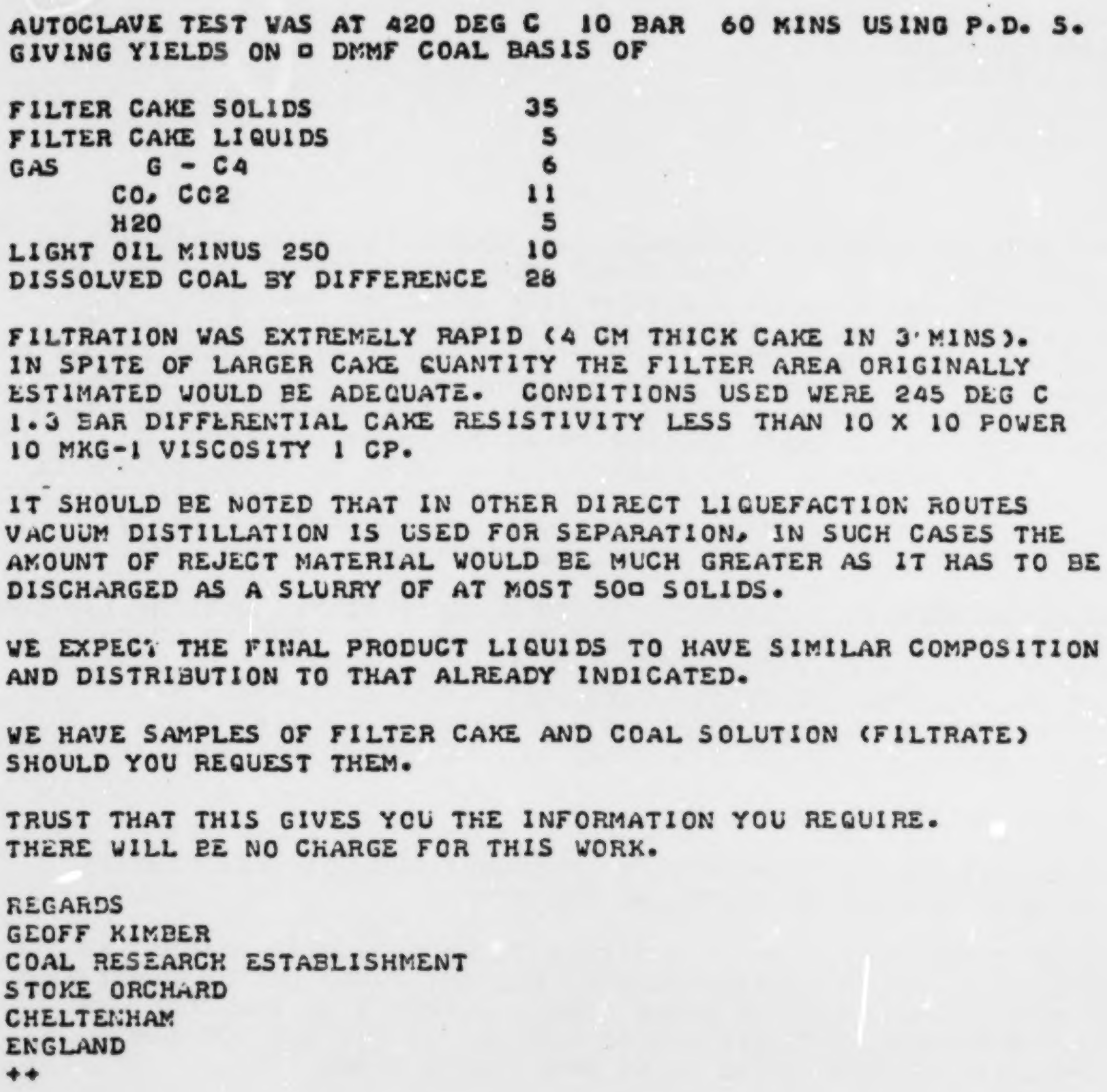
BRIEF DETAILS OF EXPERIMENTS

COAL PREPARATION PRESENTED NO PROBLEMS AND MOISTURE CONTENT OF COAL CHARGED TO AUTOCLAVE WAS 23□.

BOMB TESTS WERE PERFORMED WITH BOTH HYDROGENATED ANTHRACENE OIL (HAO) AND A PROCESS DERIVED SOLVENT (PDS) FROM A PLANT RUN WITH ANOTHER CANADIAN BROWN COAL. SLURRIES OF 33□ COAL A.R. 25□ DMMF WERE DIGESTED FOR 60 MINUTES AT 410 AND 430 DEG C. UNDISSOLVED COAL BASED ON QI DMMF BASIS WERE:

PDS	410 DEG C	36□
PDS	430 DEG C	32□
HAO	410 DEG C	32□
HAO	430 DEG C	30□

Appendix D-1 Results of Micro-autoclave Tests on Willow Bunch Lignite (Coronach #3216) Carried Out at Coal Research Establishment, National Coal Board (cont'd)



AUTOCLAVE TEST WAS AT 420 DEG C 10 BAR 60 MINS USING P.D. S. GIVING YIELDS ON □ DMMF COAL BASIS OF

FILTER CAKE SOLIDS		35
FILTER CAKE LIQUIDS		5
GAS	G - C4	6
	CO, CO2	11
	H2O	5
LIGHT OIL MINUS 250		10
DISSOLVED COAL BY DIFFERENCE		28

FILTRATION WAS EXTREMELY RAPID (4 CM THICK CAKE IN 3 MINS). IN SPITE OF LARGER CAKE QUANTITY THE FILTER AREA ORIGINALLY ESTIMATED WOULD BE ADEQUATE. CONDITIONS USED WERE 245 DEG C 1.3 BAR DIFFERENTIAL CAKE RESISTIVITY LESS THAN 10 X 10 POWER 10 MKG-1 VISCOSITY 1 CP.

IT SHOULD BE NOTED THAT IN OTHER DIRECT LIQUEFACTION ROUTES VACUUM DISTILLATION IS USED FOR SEPARATION, IN SUCH CASES THE AMOUNT OF REJECT MATERIAL WOULD BE MUCH GREATER AS IT HAS TO BE DISCHARGED AS A SLURRY OF AT MOST 50□ SOLIDS.

WE EXPECT THE FINAL PRODUCT LIQUIDS TO HAVE SIMILAR COMPOSITION AND DISTRIBUTION TO THAT ALREADY INDICATED.

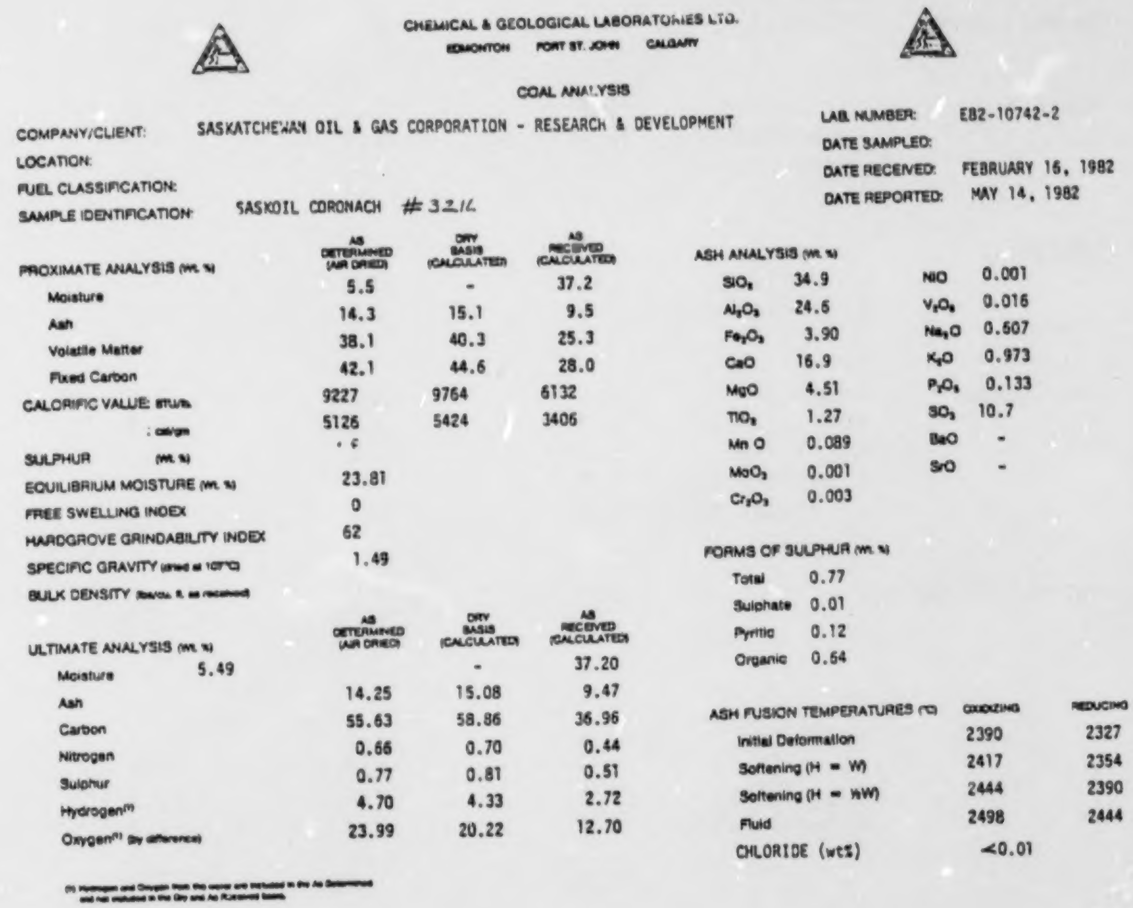
WE HAVE SAMPLES OF FILTER CAKE AND COAL SOLUTION (FILTRATE) SHOULD YOU REQUEST THEM.

TRUST THAT THIS GIVES YOU THE INFORMATION YOU REQUIRE. THERE WILL BE NO CHARGE FOR THIS WORK.

REGARDS
GEOFF KIMBER
COAL RESEARCH ESTABLISHMENT
STOKE ORCHARD
CHELTENHAM
ENGLAND
++

Source: Telex from Mr. G. Kimber, Coal Research Establishment, NCB, dated February 22, 1982.

Appendix D-2 Ultimate, Proximate and Ash Analyses of Willow Bunch Lignite (Coronach #3216)



CHEMICAL & GEOLOGICAL LABORATORIES LTD.
EDMONTON FORT ST. JOHN CALGARY

COAL ANALYSIS

COMPANY/CLIENT: SASKATCHEWAN OIL & GAS CORPORATION - RESEARCH & DEVELOPMENT
LOCATION:
FUEL CLASSIFICATION:
SAMPLE IDENTIFICATION: SASKOIL CORONACH #3216

LAB. NUMBER: E82-10742-2
DATE SAMPLED:
DATE RECEIVED: FEBRUARY 16, 1982
DATE REPORTED: MAY 14, 1982

	AS DETERMINED (AIR DRIED)	DRY BASIS (CALCULATED)	AS RECEIVED (CALCULATED)
PROXIMATE ANALYSIS (wt. %)			
Moisture	5.5	-	37.2
Ash	14.3	15.1	9.5
Volatile Matter	38.1	40.3	25.3
Fixed Carbon	42.1	44.6	28.0
CALORIFIC VALUE: BTU/lb.	9227	9764	6132
: cal/gm	5126	5424	3406
SULPHUR (wt. %)			
EQUILIBRIUM MOISTURE (wt. %)	23.81		
FREE SWELLING INDEX	0		
HARDGROVE GRINDABILITY INDEX	62		
SPECIFIC GRAVITY (dried at 107°C)	1.49		
BULK DENSITY (lbs/cu. ft. as received)			

ULTIMATE ANALYSIS (wt. %)		AS DETERMINED (AIR DRIED)	DRY BASIS (CALCULATED)	AS RECEIVED (CALCULATED)
Moisture	5.49		-	37.20
Ash		14.25	15.08	9.47
Carbon		55.63	58.86	36.96
Nitrogen		0.66	0.70	0.44
Sulphur		0.77	0.81	0.51
Hydrogen(1)		4.70	4.33	2.72
Oxygen(1) (by difference)		23.99	20.22	12.70

(1) Hydrogen and Oxygen from the water are included in the As Determined and not included in the Dry and As Received basis.

ASH ANALYSIS (wt. %)

SiO_2	34.9	NiO	0.001
Al_2O_3	24.6	V_2O_5	0.016
Fe_2O_3	3.90	Na_2O	0.607
CaO	16.9	K_2O	0.973
MgO	4.51	P_2O_5	0.133
TiO_2	1.27	SO_3	10.7
Mn O	0.089	BaO	-
MoO_3	0.001	SrO	-
Cr_2O_3	0.003		

FORMS OF SULPHUR (wt. %)

Total	0.77
Sulphate	0.01
Pyritic	0.12
Organic	0.64

ASH FUSION TEMPERATURES (°C)	OXIDIZING	REDUCING
Initial Deformation	2390	2327
Softening (H = W)	2417	2354
Softening (H = ½W)	2444	2390
Fluid	2498	2444
CHLORIDE (wt%)	<0.01	

---

* Analysis of coal sample tested by Coal Research Establishment, NCB, in the microautoclave unit of LSE process.

Appendix D-3 Petrographic Analyses of Willow Bunch (Coronach) Lignite

Maceral group	Submaceral/ Maceral	Coronach 3231	Coronach 3216	Coronach 3218
		%	%	%
Huminite	Structured	45.6	55.0	63.0
	Groundmass	18.8	16.4	15.4
	Unstructured	3.2	2.0	7.0
Liptinite	Sporinite	1.0	1.2	0.6
	Resinite	-	0.2	-
	Liptodetrinite	1.6	0.6	0.8
Inertinite	Fusinite	12.6	8.0	3.0
	Semifusinte	5.4	8.4	2.4
	Macrinite	0.2	0.2	-
	Sclerotinite	0.8	-	-
	Inertodetrinite	8.8	6.0	6.8
Mineral Matter		2.0	2.0	1.0
		100%	100%	100%
Total	Huminite	67.6	73.4	85.4
	Liptinite	2.6	2.0	1.4
	Inertinite	27.8	22.6	12.2
	Mineral Matter	2.0	2.0	1.0
		100%	100%	100%

* Sample #3216 was tested by Coal Research Establishment, NCB, in the microautoclave unit of LSE process.

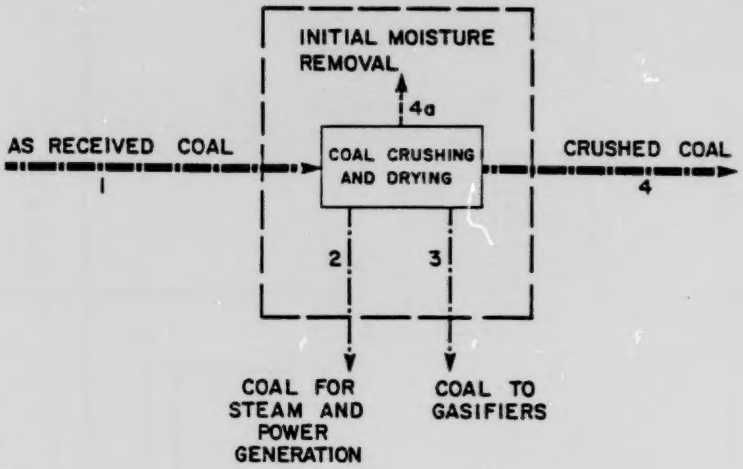
Source: Energy Research Unit, University of Regina, excerpt from the report prepared for Saskoil.

APPENDIX E

OVERALL MATERIAL BALANCE OF A CONCEPTUAL COMMERCIAL PLANT

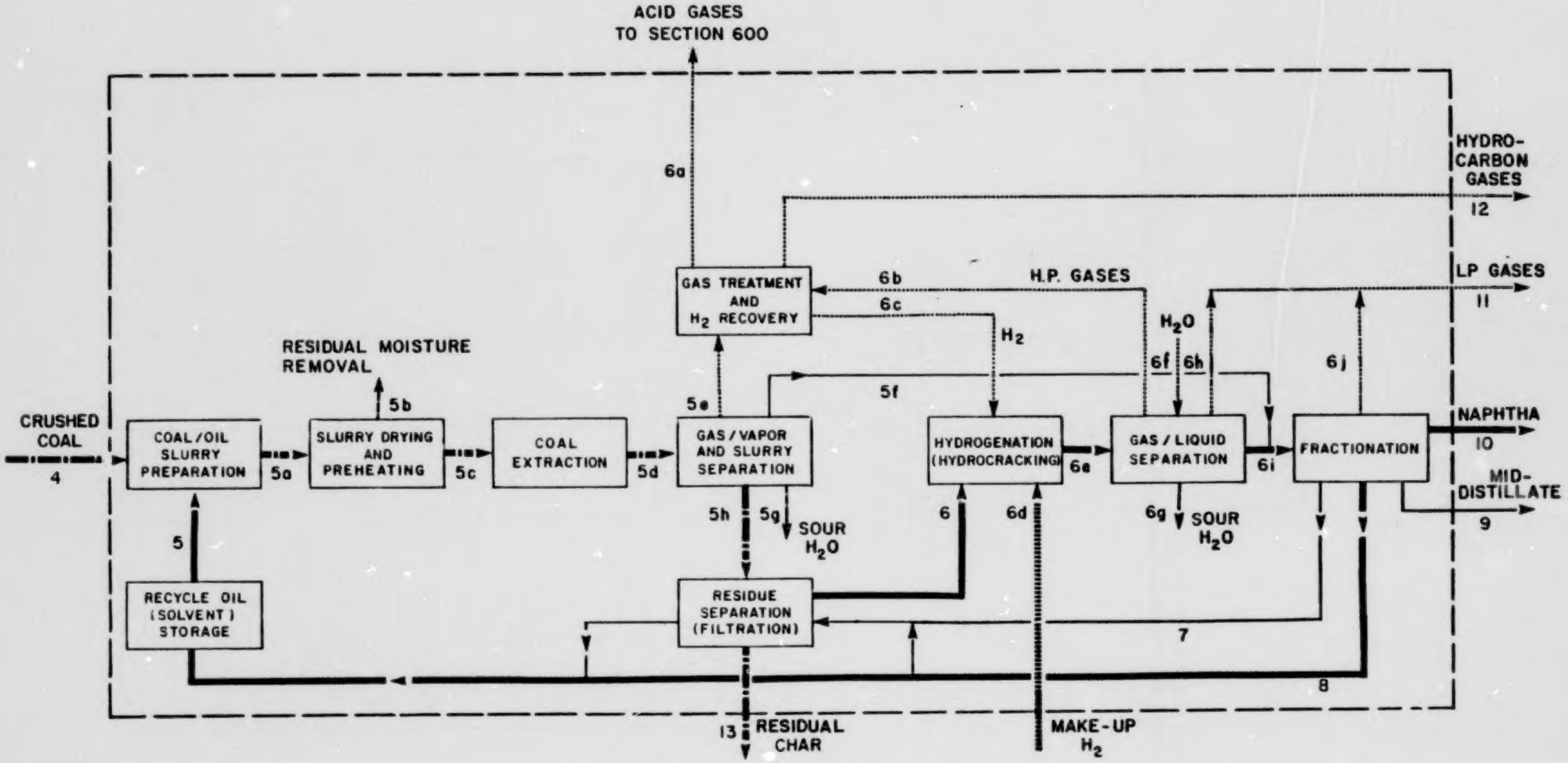
Case: 65% Extraction Efficiency

Appendix E-1 Coal Preparation (Section 100)



Stream No.	1	2	3	4	4a
Stream Identification	As-Received Coal	Coal for Steam & Power	Coal to Gasifiers	Coal to Extractor	Moisture
Components					
H_2					
H_2S					
H_2O/Steam	8750	331	66	2498	5855
NH_3					
CO					
CO_2					
C_1/C_2					
C_3/C_4					
C_5-200°C					
200-300°C					
Solvent (> 300°C)					
Extract					
Lignite (MAF)	12839	656	1001	11182	
Ash	3411	175	266	2970	
Total, t/d	25000	1162	1333	16650	5855

Appendix E-2 Coal Liquefaction (Section 200)


ACID GASES TO SECTION 600
6a
HYDRO-CARBON GASES
12
GAS TREATMENT AND H_2 RECOVERY
6b
6c
H.P. GASES
LP GASES
11
H_2
H_2O
RESIDUAL MOISTURE REMOVAL
6f
6h
6j
5b
5e
5f
CRUSHED COAL
4
COAL/OIL SLURRY PREPARATION
5a
SLURRY DRYING AND PREHEATING
5c
COAL EXTRACTION
5d
GAS/VAPOR AND SLURRY SEPARATION
HYDROGENATION (HYDROCRACKING)
6e
GAS/LIQUID SEPARATION
6i
FRACTIONATION
NAPHTHA
10
MID-DISTILLATE
9
5h
5g
SOUR H_2O
6
6d
6g
SOUR H_2O
5
RECYCLE OIL (SOLVENT) STORAGE
RESIDUE SEPARATION (FILTRATION)
7
8
13
RESIDUAL CHAR
MAKE-UP H_2

Appendix E-2 Coal Liquefaction (Section 200) (cont'd)

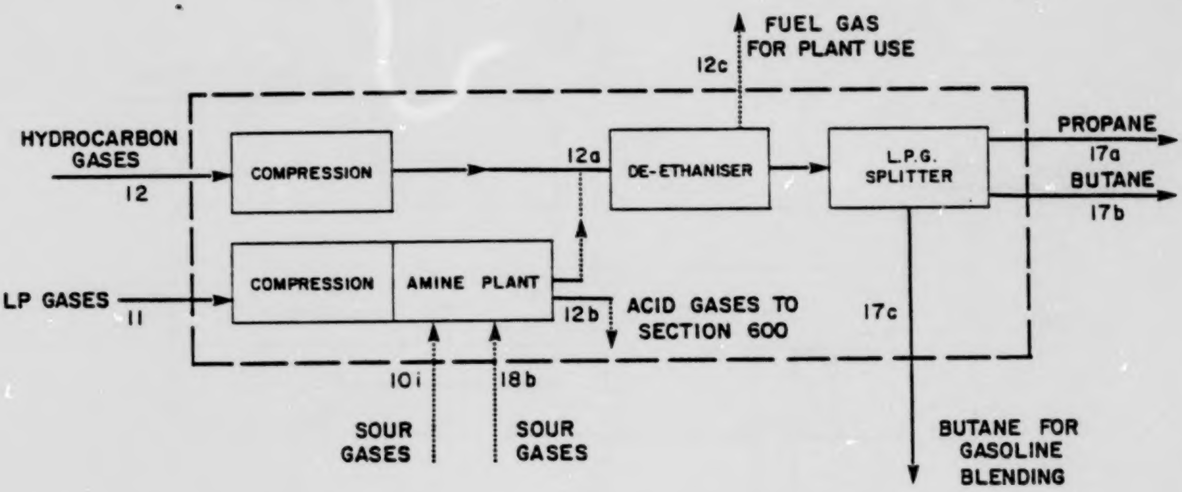
Stream No.	4	5	5a	5b	5c	5d	5e	5f	5g	5h
Stream Identification	Coal to Extractor	Solvent	Slurry	Moisture	Slurry to Extractor	Extractor Effluents	Gas	Hydro-carbon Condensate	Sour H_2O	Slurry
Components										
H_2						16	16			
H_2S						7	2		5	
H_2O/Steam	2498		2498	2165	333	1030	5		1025	
NH_3						38			38	
CO						70	70			
CO_2						575	571		4	
C_1/C_2						300	300			
C_3/C_4						239	239			
C_5-200°C						496		446		50
200-300°C						601		541		60
Solvent (> 300°C)		33328	33328		33328	33142				33142
Extract		499	499		499	4911				4911
Lignite (MAF)	11182		11182		11182	3914				3914
Ash	2970		2970		2970	2970				2970
Phenol						3			3	
Total, t/d	16650	33827	50477	2165	48312	48312	1203	987	1075	45047

Stream No.	6	6a	6b	6c	6d	6e	6f	6g
Stream Identification	Filtrate	Acid Gases	HP Gases	Recycle H_2	Make-up H_2	Hydro-Cracker Effluent	Wash Water	Sour H_2O
Components								
H_2			70	84	648	73		
H_2S		29	27			37		3
H_2O/Steam		5				729	648	1485
NH_3						18		17
CO			16	2	1	17		
CO_2		543						
C_1/C_2		11	359	24	197	497		
C_3/C_4		5				257		
C_5-200°C	50					943		
200-300°C	60					2240		
Solvent (> 300°C)	33140					33644		
Extract	4281					9		
Lignite (MAF)								
Ash								
Phenol						23		23
Total, t/d	37531	593	472	110	846	38487	648	1528

Appendix E-2 Coal Liquefaction (Section 200) (cont'd)

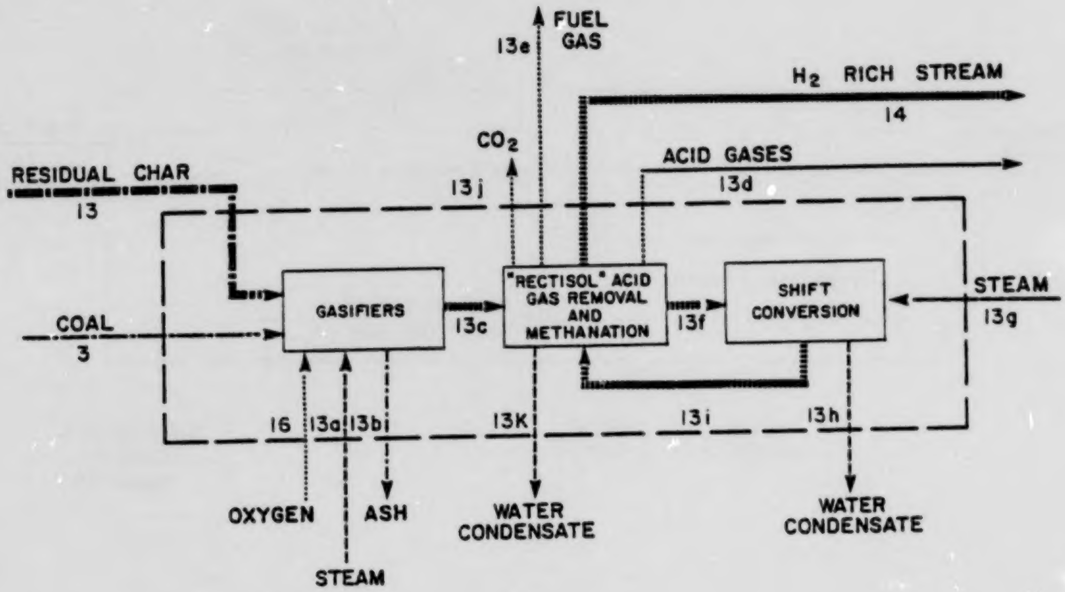
Stream No.	**6h**	**6i**	**6j**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
Stream Identification	**LP Gases**	**Hydro-crackate & solvent**	**LP Gases**	**Wash Solvent (300-370°C)**	**Recycle Solvent**	**Mid-Distillate**	**Naphtha**	**LP Gases**	**Hydro-Carbon Gases**	**Residual Gases**
Components H_2	3							3	2	
H_2S	4	3	3					7		
H_2O/Steam	4							4		
NH_3		1	1					1		
CO		1	1					1	84	
CO_2									28	
C_1/C_2	119	19	24					143	624	
C_3/C_4	70	187	191					261	234	
C_5-200°C		943					1390			
200-300°C		2240				2780				
Solvent (> 300°C)		33644		1284	32360					318
Extract		9								131
Lignite (MAF)										3914
Ash										2970
Total, t/d	200	37047	220	1284	32360	2780	1390	420	972	7333

Appendix E-3 Gas Treatment and Light-Ends Separation (Section 300)



Stream No.	10i	11	12	12a	12b	12c	17a	17b/17c	18b
Stream Identification	Sour Gases	LP Gases	Hydro-Carbon Gases	Stream to de-elthanizer	Acid Gases	Fuel Gas	Propane	Butane	Sour Gases
Components									
H_2		3	2	5		5			
H_2S	0.1	7			7				0.1
H_2O/Steam		4			4				
NH_3		1			1				
CO	0.1	1	84	84	2	84.5			0.5
CO_2			28		28				
C_1/C_2	35	143	624	869	3	869			70.6
C_3/C_4	34.2	261	234	557	2	24	291	144/98	30.0
C_5-200°C									
200-300°C									
Solvent (> 300°C)									
Extract									
Lignite (MAF)									
Ash									
Total, t/d	69.4	420	972	1515	47	983	291	144/98	101.2

Appendix E-4 Hydrogen and Fuel Gas Production (Section 400)

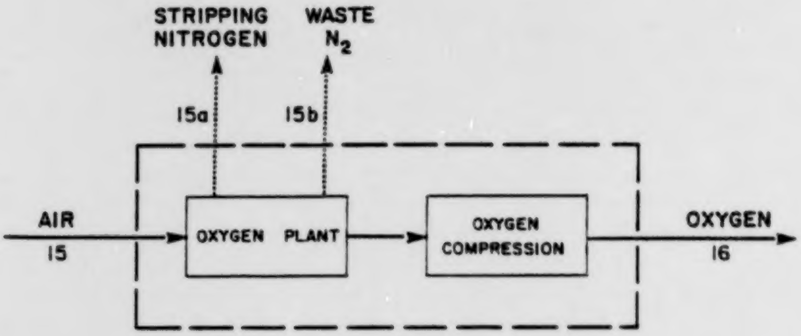


Stream No.	3	13	13a	13b	13c	13d	13e	13f	13g	
Stream Identification	Coal to Gasifiers	Residual Char	Steam	Ash	Raw Gas	Acid Gases	Fuel Gas	Syngas	Steam/ Water	
Components										
H_2					342		54	288		
H_2S					82	82				
H_2O/Steam	66		796		225	5	37	183	12384	
NH_3										
CO					8299		1366	6933		
CO_2					2327	9	401	1917		
C_1/C_2										
C_3/C_4										
C_5-200°C										
200-300°C										
Solvent (> 300°C)		318								
Extract		131								
Lignite (MAF)	1001	3914								
Ash	266	2970		3236						
N_2/Ar					120		20	100		
Total, t/d	1333	7333	796	3236	11395	96	1878	9421	12384	

Appendix E-4 Hydrogen and Fuel Gas Production (Section 400) (cont'd)

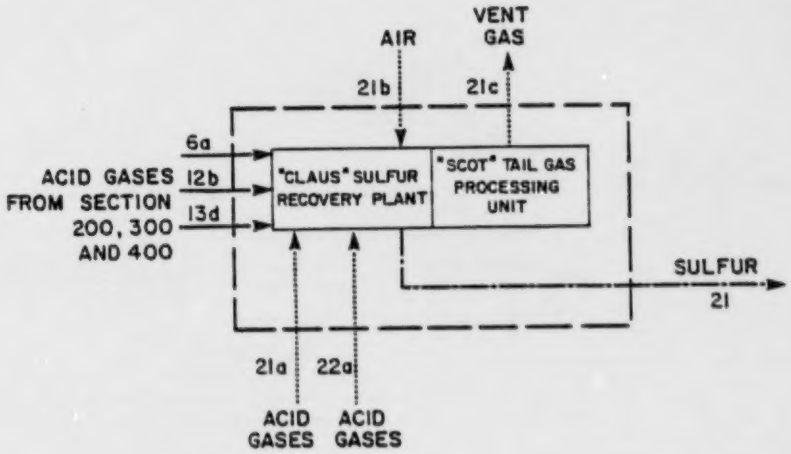
Stream No.	13h	13i	13j	13k	14	16
Stream Identification	Water Condensate	Shifted Gas	CO_2	Water Condensate	H_2-rich Stream	Oxygen
Components						
H_2		776	2		680	
H_2S				381		
H_2O/Steam	8210					
NH_3						
CO		168	27		1	
CO_2		12551	12198			
C_1/C_2					206	
C_3/C_4						
C_5-200°C						
200-300°C						
Solvent (> 300°C)						
Extract						
Lignite (MAF)						
Ash						
N_2/Ar		100	99		1	120
Oxygen						5049
Total, t/d	8210	13595	12326	381	888	5169

Appendix E-5 Oxygen Plant (Section 500)



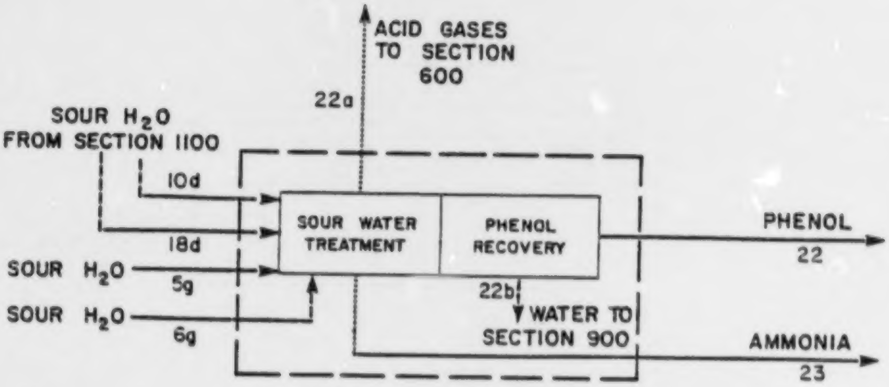
Stream No.	15	15a	15b	16
Stream Identification	Air	Stripping Nitrogen	Waste Nitrogen	Oxygen
Components				
H_2				
H_2S				
H_2O/Steam				
NH_3				
CO				
CO_2				
C_1/C_2				
C_3/C_4				
C_5-200°C				
200-300°C				
Solvent (> 300°C)				
Extract				
Lignite (MAF)				
Ash				
N_2/Ar	18665	1435	17110	120
Oxygen	5642	1	592	5049
Total, t/d	24307	1436	17702	5169

Appendix E-6 Emission Control (Section 600)



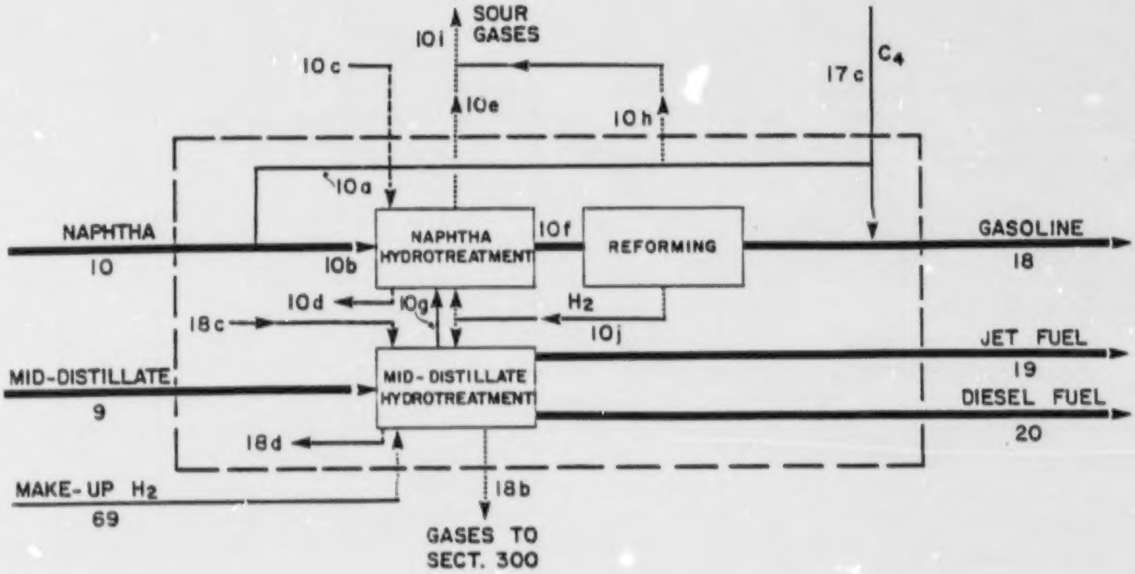
Stream No.	6a	12b	13d	21	21a	21b	21c	22a
Stream Identification	Acid Gases (Sect 200)	Acid Gases (Sect 300)	Acid Gases (Sec 400)	Sulfur	Acid Gases (Sect 900)	Air	Vent Gases	Acid Gases (Sect 700)
Components H_2								
H_2S/SO_2 (S)	29	7	82	(125)	7		0.5	8
H_2O/Steam	5	4	5		11		106	10
NH_3		1					1	
CO		2					2	
CO_2	543	28	9		28		612	4
C_1/C_2	11	3					14	
C_3/C_4	5	2					7	
C_5-200°C								
200-300°C								
Solvent (> 300°C)								
Extract								
Lignite (MAF)								
Ash								
N_2/Ar						246	246	
Oxygen						74	11	
Total, t/d	593	47	96	(125)	46	320	999	22

Appendix E-7 Effluent Control (Section 700)



Stream No.	59	69	10d	18d	22	22a	22b	23
Stream Identification	Sour Water (Sect 200)	Sour Water (Sec 200)	Sour Water (Sec 1100)	Sour Water (Sec 1100)	Phenol	Acid Gases (Sect 600)	Re-usable H_2O	Ammonia
Components								
H_2								
H_2S	5	3	0.1	0.1		8		
H_2O/Steam	1025	1485	54.6	189	6	10	2738	
NH_3	38	17	2	7				64
CO								
CO_2	4					4		
C_1/C_2								
C_3/C_4								
C_5-200°C								
200-300°C								
Solvent (> 300°C)								
Extract								
Lignite (MAF)								
Ash								
Phenol	3	23			26			
Total, t/d	1075	1528	56.7	196.1	32	22	2738	64

Appendix E-8 Product Upgrading (Section 1100)



Stream No.	9	10	17c	10a	10b	10c	10d	10e	10f	10g
Stream Identification	Mid-Distillate	Naphtha	Butane for Blending	Naphtha By-pass	Naphtha to Hydrotreater	Wash Water	Sour Water	Gases to Sect 300	Hydro-treated Naphtha	Hydro-treated Naphtha
Components										
H_2										
H_2S							0.1			
H_2O/Steam						50	54.6			
NH_3							2			
CO								0.1		
CO_2										
C_1/C_2								13.0		
C_3/C_4			144					5.2		
C_5-200°C		1390		571	819				803	58
200-300°C	2780									
Solvent (> 300°C)										
Extract										
Lignite (MAF)										
Ash										
Total, t/d	2780	1390	144	571	819	50	56.7	18.3	803	58

Appendix E-8 Product Upgrading (Section 1100) (cont'd)

Stream No.	10h	10i	10j	18	18a	18b	18c	18d	19	20
Stream Identification	Reformer Gas	Sour Gas	Re-cycle H_2	Gasoline	Make-up H_2	Sour Gas	Wash Water	Sour Water	Jet Fuel	Diesel Fuel
Components										
H_2			27		32			0.1		
H_2S		0.1				0.1	173	189		
H_2O/Steam								7		
NH_3										
CO		0.1	0.5			0.5				
CO_2										
C_1/C_2	22	35	7.3		9	70.6				
C_3, C_4	29	34.2				30.0				
C_5-200°C				1490						
200-300°C									1402	1262
Solvent (> 300°C)										
Extract										
Lignite (MAF)										
Ash										
Total, t/d	51	69.4	34.8	1490	41	101.2	173	196.1	1402	1262

APPENDIX F

REGULATORY ENVIRONMENT CONTROL ACTS
PROVINCE OF SASKATCHEWAN

APPENDIX F: Regulatory Environmental Control Act - Province of Saskatchewan

The proposed lignite liquefaction plant in the province of Saskatchewan will have to comply with the applicable regulations under the following acts:

1. THE AIR POLLUTION CONTROL ACT

This act requires that permits are secured for operating fuel-burning equipment which certainly would pertain to any of the lignite liquefaction route (direct or indirect) under consideration. The Air Pollution Control Regulations contains ambient air quality standards which are summarized in Table F-1 for the air pollutants of suspended particulates, sulfur dioxide and nitrogen dioxide.

2. **THE WATER RESOURCES MANAGEMENT ACT**

The act requires that construction and operational permits are obtained in respect of proposed water and sewage works or polluted effluent discharges. The Water Pollution Control and Waterworks Regulations specify the permit application requirements. Guidelines regarding the design of water and sewage works are given in separate documents, which are exclusively pertaining to municipal works. Regarding liquid effluents from the lignite liquefaction plant under consideration, it is to be expected that project specific water pollution requirements would be defined by Saskatchewan Environment during the course of a permit application. Some general policy statements and effluent discharge objectives are given in the Water Quality Objectives issued by Saskatchewan Environment. This document includes several numerical objectives which are summarized in Tables F-2 through F-5.

3. **THE WATER RIGHTS ACT**

The act requires that approvals are secured in terms of surface water use, obstruction, diversions or storage. These provisions would possibly apply to the cooling system water supply chosen for any of the direct or indirect lignite liquefaction plant.

4. **THE GROUND WATER CONSERVATION ACT**

The act requires that permits and licenses be secured prior to the exploration for and use of ground waters respectively. The Ground Water Regulations specify the requirements for obtaining a permit or license. The legislation also contains provisions for prevention of ground water pollution. The possible use of ground water in the direct or indirect lignite liquefaction plants as well as the ground water dewatering systems for the proposed mine would require several approvals under this act.

5. THE PUBLIC HEALTH ACT

The act mostly contains provisions regarding hygienic matters. The Waste Management Regulations of 1972 may however apply to the disposal of ash and slag rejects from the lignite liquefaction plant. The regulations stipulate that a permit be secured in respect of the location, operation and maintenance of solid waste disposal systems. The regulations also contain provisions for cooperation between the Public Health and Environment Departments in terms of setting directives for the disposal of toxic wastes or substances.

Table F-1 Ambient Air Quality Standards - Province of Saskatchewan

Air Contaminant	Unit	Average Concentration for Applicable Time Period				
		1 hour	8 hours	24 hours	30 days	1 year
Suspended Particulates	ug/m^3	- - -	- - -	120	- - -	70 *
Settable Particulates	mg/cm^2	- - -	- - -	- - -	2	- - -
Soil Index	COH units	- - -	- - -	1.5	- - -	- - -
Potash	mgK+/cm^2 mgCl$^-$/cm^2	- - - - - -	- - - - - -	- - - - - -	0.15 or 0.15	- - - - - -
Sulphur Dioxide ***	ug/m^3 ppm (volumetric)	450 0.17	- - - - - -	150 0.06	- - - - - -	30 * 0.01
Sulphation	mg/100 cm^2	15	- - -	- - -	30	- - -
Carbon Monoxide	mg/m^3 ppm (volumetric)	15 13	6 5	- - - - - -	- - - - - -	- - - - - -
Oxidents (Ozone)	ug/m^3 ppm (volumetric)	100 0.5	- - - - - -	30 0.015	- - - - - -	- - - - - -
Nitrogen Dioxide	ug/m^3 ppm (volumetric)	400 0.2	- - - - - -	200 0.1	- - - - - -	100 ** 0.05

* Geometric means.
** Arithmetic means.
*** Montana ambient standards for SO_2: 0.25 ppm (1 hr.), 0.10 ppm (24 hr.), 0.02 ppm (annual).

NOTES: No hydrogen sulfide gas can be released to the atmosphere from any industrial source, fuel-burning equipment or incinerator. Opacity shall be less than 40% or not more shady than No. 2 on the Ringelmann Chart; discharges during start-up or shut-down, in excess of 40% opacity are permitted for a duration of less than 5 minutes.

SOURCE: Environment Saskatchewan, Air Pollution Control Regulations, 211/75, 1975.

Table F-2 Saskatchewan Surface Water Quality Objectives*

Parameter	Objectives
Bacteriology (Coliform Group)	- In water to be withdrawn for treatment and distribution as a potable supply or used for outdoor recreation other than direct contact, at least 90 percent of the samples (not less than five samples in any consecutive 30-day period) should have a total coliform density of less than 5,000 per 100 mL and a fecal coliform density of less than 1,000 per 100 mL. - In waters used for direct contact recreation or vegetable crop irrigation the geometric mean of not less than five samples taken over not more than a 30-day period should not exceed 1,000 per 100 mL total coliforms, nor exceed these numbers in more than 20 percent of the samples examined during any month, nor exceed 2,400 per 100 mL total coliforms on any day.
Physical Quality	
Dissolved Oxygen	- A minimum of 5 mg/L at any time.
Biochemical Oxygen Demand	- Dependent on the assimilative capacity of the receiving water, the BOD must not exceed a limit which would create a dissolved oxygen content of less than 5 mg/L.
Suspended Solids	- Not to be increased by more than 10 mg/L over background value.
pH	- To be in the range of 6.5 to 8.5 pH units but not altered by more than 0.5 pH units from background value.
Temperature	- Not to be exceeded by more than 3°C above ambient water temperature.
Odour	- The cold (20°C) threshold odour number not to exceed eight.
Colour	- Not to be increased more than 30 colour units above natural value.
Turbidity	- Not exceed more than 25 Jackson units over natural turbidity.

* Apply to surface waters, except in areas of close proximity to outfalls.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table F-2 Saskatchewan Surface Water Quality Objectives (cont'd)

Parameter	Objectives (Maximum Concentrations)
	mg/L
Organic Chemical	
Carbon Chloroform Extract (CCE) (includes Carbon Alcohol Extract)	0.2
Methyl Mercaptan	0.05
Methylene Blue Active Substances	0.5
Oil and Grease	Substantially absent, no irridescent sheen
Phenolics	0.005
Resin Acids	0.1
Inorganic Chemical	
Boron	0.5
Copper	0.02
Fluoride	1.5
Iron	0.3
Manganese	0.05
Nitrogen (Total Inorganic and Organic)*	1.0
Phosphorous as PO_4 (Total Inorganic and Organic)*	0.15
Sodium (as percent of cations)	30 - 75
Sulphide	0.05
Zinc	0.05
Toxic Chemicals	
Arsenic	0.01
Barium	1.0
Cadmium	0.01
Chromium	0.05
Cyanide	0.01
Lead	0.05
Mercury	0.0001
Selenium	0.01
Silver	0.05

Notes: The predominant cations of sodium, calcium and magnesium and anions of sulphate, chloride and bicarbonate are too variable in the natural water quality state to attempt to define limits. Nevertheless, in order to prevent impairment of water quality, where effluents containing these ions are discharged to a water body the ermissible concentration will be determined by the Department in accordance with existing quality and use. Unless specified, parameter concentrations refer to "total" measurements.

* These objectives are presently under study and may require adjusting according to naturally occurring concentrations or conditions.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table F-2 Saskatchewan Surface Water Quality Objectives (cont'd)

Parameter	Objectives
Pesticides	- To provide reasonably safe concentrations of these materials in receiving waters an application shall not exceed 1/100 of the 48-hour Tl_m. Consideration must also be given to sublethal effects such as taste and odour generation.
Radioactivity	- Gross Beta not to exceed 1,000 pCi/L. - Radium 226 not to exceed 3 pCi/L. - Strontium 90 not to exceed 10 pCi/L.
Unspecified Substances	- Substances not specified herein should not exceed values which are considered to be deleterious for the most critical use as established by the Department.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table F-3 Saskatchewan Municipal Drinking Water Quality Objectives

Parameter	Objectives
Bacteriology	
Total Coliform	- At least 90 percent of the samples in any consecutive 30 day period should be negative for total coliform organisms and no one sample should have a Most Probably Number index of greater than 10 per 100 mL or 30 per 100 mL by membrane filter. Properly operated municipal waterworks should be free of coliform bacteria.
Nuisance Biological Organism	- Biological organisms in concentrations which may produce objectionable colour, taste, odour and turbidity, or which may release toxid metabolites, or which may harbour pathogens are undesirable in drinking water and should be kept below such concentrations as to prevent any undesirable effects.
Radioactivity	
Gross Alpha Activity	- Not to exceed 10 pCi/L.
Strontium 90	- Not to exceed 10 pCi/L.
Radium 226	- Not to exceed 3 pCi/L.
Gross Beta Activity	- Not to exceed 1000 pCi/L (in absence of Alpha).
Other Specific Nuclides	- Not to exceed 0.1 of the maximum permissible concentration in water adopted by the International Commission on Radiological Protection.
Physical Quality	
	- Water should not contain impurities that would be offensive to the sense of sight, smell or taste.
Colour	- Maximum of 15 units.
Threshold Odor Number	- Maximum of 3 units.
Turbidity	- Maximum of 5 units.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table F-3 Saskatchewan Municipal Drinking Water Quality Objectives

Parameter	Objectives (Maximum Concentrations) mg/L
Chemical Quality	
Alkalinity ($CaCO_3$)	500*
Chloride	250
Copper	1.0
Fluoride	1.5
Iron	0.3
Hardness ($CaCO_3$)	800*
Magnesium	200*
(Magnesium & Sodium) and Sulfate	1,000*
Manganese	0.05
Methylene Blue Active Substances	0.5
Organics (carbon chloroform and carbon alcohol extractibles)	0.2
Phenolics	0.001*
Sodium	300*
Sulphate	500
Total Dissolved Solids	1,500*
Zinc	5.0
pH (permissable range)	7.0 - 9.5
Toxic Chemical	
Arsenic	0.01
Barium	1.0
Boron	5.0
Cadmium	0.01
Chromium	0.05
Cyanide	0.01
Lead	0.05
Nitrates (including nitrites) as NO_3	40*
Selenium	0.01
Silver	0.05
Biocides**	absent

* These values have been modified from those of the Canadian Drinking Water Standard and Objectives, 1968.

** In general biocides should be virtually absent or not detectable in any drinking waters. The presence of any biocide would require a special investigation and study.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table F-4 Saskatchewan Guidelines for the Chemical Water Quality Respecting Private Water Supplies

Constituent	Range of Concentration Satisfactory Quality	Poor Quality	Not recommended for Consumption	Unsuitable for Use
	mg/L	mg/L	mg/L	
Total Dissolved Solids	100 - 1,500	1,500 - 3,000	3,000 - 4,000	GT 4,000
Total Hardness	50 - 500	500 - 1,000	1,000 - 2,000	GT 2,000
Total Alkalinity ($CaCO_3$)	50 - 500	500 - 1,000	1,000 - 1,500	GT 1,500
Chloride	LT 250	250 - 500	500 - 1,000	GT 1,000
Sodium	LT 300	300 - 500	500 - 1,000	GT 1,000
Sulphate	LT 400	400 - 800	800 - 1,200	GT 1,200
Nitrate	LT 40	40 - 300	GT 300	- - - - -
Iron	LT 0.3	0.3 - 1.0	- - - - - - - -	- - - - -
Manganese	LT 0.05	0.05 - 0.5	- - - - - - - -	- - - - -
pH (units)	7.0 - 9.5	6.0 - 7.0 and 9.5 - 10.0	- - - - - - - -	LT 5.5 and GT 10.5

Abreviations, LT: Less than, GT: Greater than.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table F-5 Saskatchewan Guidelines for Sewage Treatment

Parameter	Percent Removal	Effluent Value
	%	mg/L*
Primary Treatment (including anaerobic lagoons)		
BOD_5 @ 20°C	25 - 45	100 - 150
Coliform Bacteria (per 100 mL)	40 - 60	25×10^6
Methylene Blue Active Substances	Nil	- - - - -
Settable Solids	70 - 90	0.5 - 1.0
Suspended Solids	33 - 65	100 - 150
Secondary Treatment (including aerated lagoons)		
Ammonia Nitrogen	- - - - -	10 - 20
BOD_5 @ 20°C	85 - 95	15 - 30
Coliform Bacteria (per 100 mL)	99.5- 99.9	20,000 - 2,000,000
Methylene Blue Active Substances	80 - 95	0.2 - 0.8
Dissolved Oxygen	- - - - -	3 - 5
Oil and Grease (ether solubles)**	- - - - -	15 - 25
ph (units)	- - - - -	5.5 - 10.5
Phenolics **	- - - - -	0.02 - 0.04
Suspended Solids	85 - 96	15 - 30
Phosphorus (PO_4)	- - - - -	15 - 30

Specific Guidelines for Conventional Aerobic Lagoons

Parameter	Percent Removal		Effluent Value
	Winter	Summer	Summer
	%	%	mg/L*
BOD_5 @ 20°C	25-65	85-95	15 - 30
Coliform Bacteria (per 100 mL)	GT90	GT99.5	20,000 - 200,000
Methylene Blue Active Substances	- - -	60-90	0.2 - 1.5
Suspended Solids	75-90	70-90	15 - 100***
Algae (volumetric in ppm)	- - -	- - -	150 - 300
Phosphorus (PO_4)	- - -	- - -	5 - 20

Abreviations: GT - greater than.

* Values in mg/L unless otherwise indicated for the parameter.
** Applied to any effluent regardless of treatment.
*** Includes algae.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

APPENDIX G

ECONOMICAL AND FINANCIAL ASSUMPTIONS AND CRITERIA

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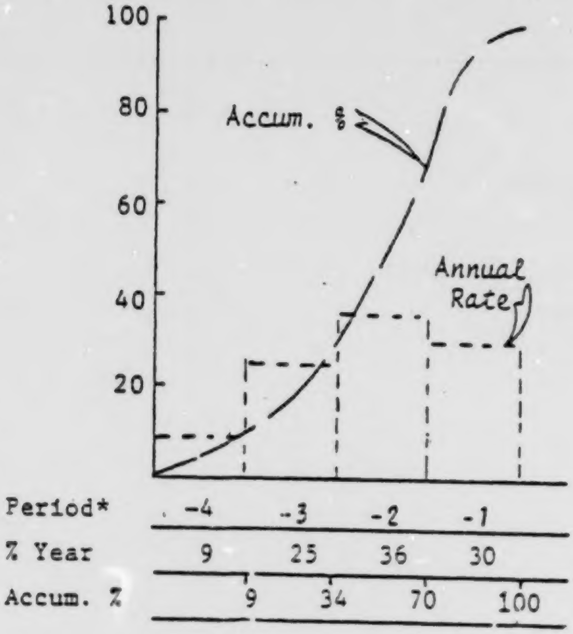
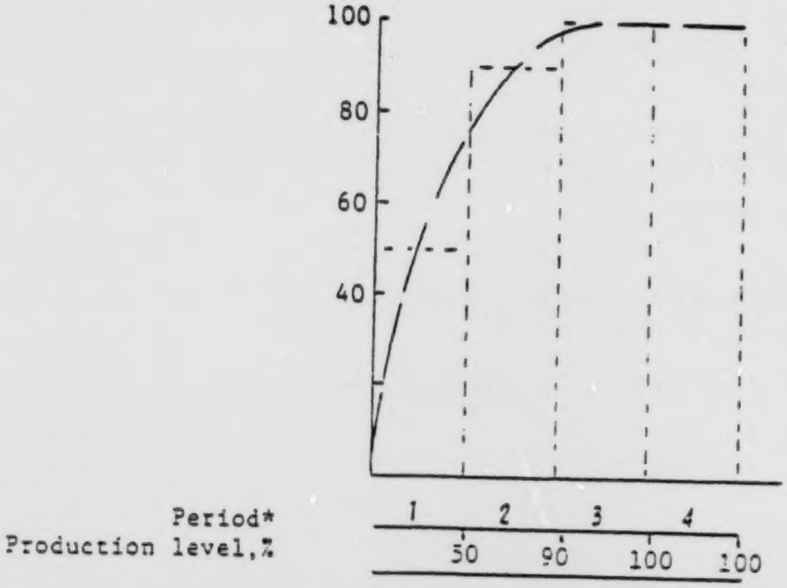
Appendix G-1: Capital Investment Schedule (Spending Schedule)



Appendix G-2 Production Level (Percentage of Annual Production Rate)



With reference to plant start-up date.

Source: B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979

Appendix G-3: Canada/U.S. Material Price Differences

	U.S.A.	CANADA	VARIANCE
Concrete Mix, U.S. $/cwt	47.90	50.12	4.6%
Standard Structures, U.S. $/cwt			
Wide Flanges	26.20	30.95	18.13%
I-Beams	27.70	33.95	22.6%
Piping, U.S. $/ft			
Iron	4.51	6.44	42.8
Copper	0.81	0.55	32.1
Cables/Wires			10%

Assuming the material distribution of:

- 30% concrete
- 15% steel (50% wide flanges, 50% I-beams)
- 45% piping (80% iron, 20% copper)
- 10% cables and wires

the "Material Difference Factor" of 1.1592 is obtained.

Therefore, "Material plus Currency Differential Factor" is 1.1592 x 1.19
= 1.379

Source: Engineering News Record, Jan. 7, 1982; Jan. 14, 1982.

Appendix G-4: Conversion of Manhours to Saskatchewan (Productivity Adjustment Factor)

A review of labour productivity based on petrochemical projects indicates that a multiplier of 1.25 be applied for conversion of U.S. Gulf Coast manhours to equivalent Canadian manhours.

For adjustment to Saskatchewan, the following two items must also be considered:

Unique Condition Factor

Projects of this size will generally experience during the construction period several occurances whereby labour will interpret certain management decisions as being an affront to their integrity and pride. It will be assumed that these types of non-strike situations will lead to a cumulative loss of one working day per month during the construction period.

Factor = one day/20 working days per month = 0.05

Climatic Factor

The Saskatchewan Region experiences considerably more sub-zero temperatures and the duration of the winter season is generally about two months longer than the Sarnia Region. Therefore, an allowance of 20 days per winter will be made for adverse working conditions due to weather.

$$\text{Factor} = \frac{\text{20 days x 4 winters}}{\text{48 months x 20 days/winter}} \quad + 0.08$$

Thus, the "Overall Productivity Adjustment Factor" is

1.25 x 1.05 x 1.08 = 1.417 = 1.42

Appendix G-5: Process Equipment and Plants Price Indexes

Year	Index	% Increase
1977	181.9	7.53
1978	195.6	10.70
1979	216.6	10.90
1980	240.2	11.25
1981	267.2	11.00
1982	-	

* Assumed that similar % incease will continue in 1982.

Source: Construction Price Statistics, "Chemical and Mineral Process Plant Price Indexes", Table 19, November 1981

Appendix G-6: Typical Manpower Distribution and Average Salary Scale

	% of Total Manpower *	Average Salary $ Month **
Operating Labour (OL)	34	1575
Operation Supervisors (20% of OL)	7	2500
Maintenance Labour (ML)	42	1875
Maintenance Supervision (20% of ML)	8.5	2500
Plant Management and Administrative Support	8.5	2700

Sources:

* K.A. Rogers and R.F. Hill, ESCOE, DOE Report No. Fe-2468-51 July 1979

** Values reported in SPC Study, (Job No. C8093) "Coronach Coal Gasification Study", Volume II, October 1977, have been adjusted to mid-1982 dollars using salary index in Appendix G-7.

Appendix G-7: Salary Increases for Operating Costs Calculation (Location: Saskatoon)

Year	Index	% Increase
1976	204.0	10.25
1977	224.9	5.47
1978	237.2	6.62
1979	252.9	9.25
1980	276.3	7.38
1981	296.7	9.38*
1982*	-	

* Increase for 1982 assume as 1981 + 2%

Source: Construction Price Statistics, "Union Wage Rate Index", Table 3, November 1981

Volume

3

STUDY RELATING TO THE TECHNICAL AND ECONOMIC FEASIBILITY OF LIGNITE LIQUEFACTION IN SASKATCHEWAN

Volume III

ASSESSMENT OF ADVANCED INDIRECT COAL LIQUEFACTION PROCESSES CONSIDERED FOR COMMERCIALIZATION BY 1990

Final Report
by

TECHNICAL DEVELOPMENT DEPARTMENT, SNC Inc.

for

SASKATCHEWAN OIL AND GAS CORPORATION

and

CANMET, ENERGY, MINES AND RESOURCES, CANADA

October, 1982
Montreal

REQ. NO. 2-9203

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CONTRACT REPORT NO. OSQ82-00269

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J.A. Mikhlin, Eng, Ph.D.

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ANNEX

A Review and Evaluation of Indirect Liquefaction Process with Following Gasifiers as a part of Saskoil Coal Liquefaction Study, Foster Wheeler Report to SNC Inc, F.W. Job No. B28-8011, Nov. 1981

- Lurgi Dry Bottoms
- BGC/L Slagging Bottoms
- Koppers-Totzek
- Winkler
- Texaco

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INTRODUCTION

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INTRODUCTION

This study was carried out by the Coal Conversion Group, Technical Development Department of SNC Inc. to evaluate various coal liquefaction processing alternatives for two designated Saskatchewan lignites, i.e., from the Estevan and Willow Bunch deposits. The processes evaluated were subdivided into two categories.

- processes considered for commercialization by 1990, and
- processes considered for commercialization around 1995.

The scope of the work for the advanced processes (commercialization by 1990) included review and evaluation of direct and indirect coal liquefaction and pyrolysis processes listed in Table 1. The evaluation was made in two stages. Stage I comprises technical assessment, prescreening and ranking of the processes for both Estevan and Willow Bunch lignites. Stage II covers techno-economic assessment of processes selected in Stage I but for Willow Bunch lignite only. The ultimate objective is to determine a promising process(es) for Saskatchewan lignites.

The scope of the work in the second category (commercialization around 1995) is to ascertain state-of-the-art technology and review update of the processes and concepts listed in Table 2. A summary of the results is presented in Volume I.

In addition, an independent experimental program was carried out concurrently with this study to determine comparative reactivity of Estevan and Willow Bunch (Coronach) lignite. These tests were made by the Chemical Technology Division of B.C. Research Laboratories and the results are presented in a separate report, "Liquefaction and Pyrolysis Tests on Samples of Estevan and Coronach Lignites", Project Report No. 2-02-691, March 1982. A summary of the results is presented in Volume I.

Table 1 Processes Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization by 1990

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
H-Coal	Hydrocarbon Research Inc.	U.S.A.
RAG/Veba	Ruhrkohle AG and Veba Oel AG	F.R.G.
Rheinbraun	Rheinische Braunkohlenwerke AG	F.R.G.
Exxon Donor Solvent (EDS)	Exxon Research and Engineering Company	U.S.A.
Liquid Solvent Extraction (LSE)	National Coal Board, Coal Research Establishment	U.K.
	COAL PYROLYSIS	
Lurgi-Ruhrgas	Lurgi Mineralöltechnik GmbH and Ruhrgas AG	F.R.G.
	INDIRECT COAL LIQUEFACTION	
SASOL-Type Fischer-Tropsch	SASOL Ltd.	South Africa
Mobil Methanol-to-Gasoline (M-MTG)	Mobil Research and Development Corporation	U.S.A.
	Gasifiers for Production of Synthesis Gas	
Lurgi (dry bottom)	Lurgi Mineralöltechnik AG	F.R.G.
British Gas/Lurgi (slagging)	British Gas Corporation and Lurgi Mineralöltechnik AG	U.K. F.R.G.
Koppers-Totzek	Krupp-Koppers GmbH	F.R.G.
Texaco	Texaco Research and Development Corporation	U.S.A.
Winkler (high temperature)	Rheinische Braunkohlenwerke AG	F.R.G.
Westinghouse	Westinghouse Electric Corporation	U.S.A.

Table 2 Processes and Concepts Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization around 1995

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
Zinc Halide	Consolidation Coal Co.	U.S.A.
Dow Coal Liquefaction	Dow Chemicals Co.	U.S.A.
Consol Synthetic Fuel (CFF)	Consolidation Coal Co.	U.S.A.
Solvent Refined Coal (SRC-1)	International Coal Refining Co.	U.S.A.
Solvent Refined Lignite (SRL)	University of North Dakota	U.S.A.
CO-Steam	Grand Forks Energy Technology Center	U.S.A.
Two-Step Liquefaction	Cities Service Co. C-E Lummus Co.	U.S.A.
	COAL PYROLYSIS	
Occidental Pyrolysis	Occidental Research Corporation	U.S.A.
	INDIRECT COAL LIQUEFACTION	
	Gasifiers for Production of Synthesis Gas	
Shell Gasifier	Shell International Petroleum	The Netherlands
U-Gas	Institute of Gas Technology	U.S.A.
	CO-PROCESSING	
Coil	Hydrocarbon Research Inc.	U.S.A.
Solvolysis	Mitsubishi Heavy Industries Ltd.	Japan
Veba Coal/Oel Co-processing	Veba Oel AG	F.R.G.

OBJECTIVES OF THE STUDY

The main objectives of the study for advanced processes considered for commercialization by 1990 were identified as follows:

- to establish technical criteria for process evaluation and prescreening in Stage I and techno-economic criteria for process selection in Stage II.

Stage I, Process Evaluation and Prescreening

- to determine state-of-the-art technology and predict behavior of Estevan and Willow Bunch lignite in the processes based on available data and previous experience of the process developers with similar coals.

- prescreening and ranking of the process for further techno-economic assessment.

Stage II, Techno-Economic Assessment of Preselected Processes

- techno-economic assessment and comparison of pre-selected processes using Willow Bunch lignite as feedstock.

- selection of the promising process(es) for liquefaction of Willow Bunch lignite to produce synfuels substituting conventional transportation fuels.

- to determine constraints toward commercialization for selected process(es).

TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION

Technical and economic criteria approved by the Saskatchewan Coal Liquefaction Technical Committee (the Study Committee) for direct and indirect liquefaction and pyrolysis processes are presented in detail in Volume I, however, a brief outline is given below.

Feedstock Specifications

Two Saskatchewan lignites from the Estevan and Willow Bunch deposits were designated as feedstocks for "prescreening" of coal liquefaction and pyrolysis processes (Stage I of the Study). Typical ultimate, proximate and petrographic analysis of these lignites are summarized in Appendix A. Both coals have high moisture and ash content. While average moisture content is the same for both coals, the ash content of Willow Bunch lignite and inertinite content are higher.

For techno-economic assessment of prescreened processes (Stage II of the Study), Willow Bunch lignite was designated as the feedstock. Detailed analyses of Willow Bunch lignite showing the "best" and "worst" coal properties are given in Appendix C.

TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION (Cont'd)

Product Criteria

For prescreening of the processes (Stage I of the Study) the product criteria are:

- synthetic crude (naphtha, middle oil, heavy oil) via "direct" coal liquefaction.
- synthetic crude (naphtha, middle oil, heavy oil) via pyrolysis.
- gasoline as the main product via "indirect" coal liquefacation.
- pure methanol (99.5% purity) as transportation fuel or blendstock for gasoline via indirect coal liquefaction.

In order to allow a consistent comparison of the techno-economic assessments of preselected processes (Stage II of the Study), transportation fuels such as gasoline, diesel and jet fuel are considered as desired end products for both "direct" and "indirect" coal liquefaction processes. However, emphasis is given to production of gasoline. Light hydrocarbons (C_1-C_4) are considered as co-products.

Overall and Liquid Product Efficiency

In this study, a modified version of efficiency widely used for energy-related processes has been considered. The modified definition of "overall" product efficiency is the total gross-calorific-value (GCV) of all products divided by the total energy input of coal feed and electric power. For processes, where electricity is required from an outside source, a thermal equivalent factor of 10,000 Btu/kWh (10,550 kJ/kWh) as suggested by ESCOE*, rather than the true thermodynamic equivalent factor of 3413 Btu/kWh, has been used to allow realistic conversion loss for electric power generation.

Since the objective of the study is to investigate coal liquefaction processes for the production of transportation fuels, the "liquid" product efficiency is also calculated. Here, the liquid product refers to distillate product in the temperature range of C_5- 525°C.

Overall Plant Configuration, Plant Size and Battery Limits

All processes are considered as grass roots facilities. The plant comprises self-sufficient, integrated operation requiring coal, water and electricity as major raw material and process inputs.

The as-received coal supplied to the plant is assumed to be Run-of-Mine (ROM). No pretreatment/beneficiation of coal to reduce moisture and/or ash (minerals) contents is considered. Development of coal mine, water supply, gas delivery pipelines and booster stations, product transportation facilities, etc., are also not considered.

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* B.C. McBeath, The Engineering Society Commission on Energy Inc., "Guidelines for Coal Conversion Processes", DOE Report No.FE-2468-44, April 1979.

TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION (cont'd)

Plant Size and Capacity Factor : The conceptual, commercial plant will be based on a feedrate of 25,000 metric tons (tonnes) of as-received coal supplied to the plant. A capacity factor of 0.85, i.e., 310 working days per year is assumed.

Plant Location and Coal Delivery : The plant is assumed to be located within 16 to 24 km (10 to 15 miles) radius of the coal mines. No major site development other than leveling or grubbing has been assumed. Coal is assumed to be delivered to the plant by unit railcars. If, however, the plant site is within 3.0 km (2 miles) radius, transportation by trucks or conveyors may be economical.

Ash/Slag Disposal : Ash/slag is transported back to the mines for disposal.

Coal Fines : Coal fines may be burned for steam and power generation. Excess coal fines will be returned to the mines without any credits.

Water Availability : Water is assumed to be available to the plant site from Diefenbaker Lake through a canal to be built. Supply of water will also depend on Provincial Government priorities. It is further assumed that the cost of water will be the same as that for water presently supplied by the Saskatoon South-East Water Supply canal from Diefenbaker Lake to the potash mines and the communities along its route.

TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION (cont'd)

Steam and Power Availability : Steam is generated on site by recovering process heat and burning coal. Power is also generated onsite. Additional electricity, if required, will be purchased from the utility network and a conversion factor of 10,000 Btu/kWh (10,500 kJ/kWh) will be applied for thermal efficiency calculations.

Environmental Aspects

The conceptual plant design will meet all existing local, provincial and federal environmental regulations for particulates, liquid and gaseous effluents.

Economic Criteria

Preliminary or budget-type capital cost estimate, operating, maintenance and product cost evaluation is carried out for each of the "preselected" coal liquefaction processes only (Stage II of the Study).

Each process is evaluated as a integrated, self-sufficient, grass-roots facility processing 25,000 tonnes per day of as-received Willow Bunch lignite supplied to the plant. The plant is assumed to be located within 16 km (10 miles) of the coal mines in Saskatchewan. The service factor is assumed to be 85%, i.e., 310 operating days per year.

TECHNICAL AND ECONOMIC CRITERIA FOR PROCESS COMPARISON AND SELECTION (cont'd)

All costs are calculated in "constant" mid-1982 Canadian dollars and no forward escalation beyond mid-1982 included. Cost of capital is assumed to be 10% per annum. For 100% debt financing this value is equivalent to debt* cost in "real" terms. For 100% equity financing, it may be considered as return on equity before taxes.

Product costs for preselected coal liquefaction processes will be calculated on "Energy Value" and "Product Value" bases. The first method is based on the heating value only whereas the second also considers the quality of the individual product. The latter approach, Product Value method, was developed by The Engineering Society Commission on Energy Inc. (ESCOE)** and the objective is to allow relative comparison and ranking of the energy-related processes.

Byproduct credits have been taken considering \$65/t of sulfur, \$300/t of NH_3, and \$250/t of mixed phenols.

A large quantity of CO_2 is produced which has a potential for utilization in enhanced oil recovery operation. No credits are taken for CO_2 byproduct, since the market for such use is uncertain.

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* for calculations on "constant" dollar basis, debt cost is expressed in "real" terms, i.e., prevailing interest rates minus inflation.

** K.A. Roger and R.F. Hill, "Coal Conversion Comparison", DOE Report FE-2468-51, July 1979.

STUDY REPORT ORGANIZATION

The study report comprises of five volume:

Volume I - Summary and Process Comparison

Volume II - Assessment of Advanced Direct Coal Liquefaction and Pyrolysis Processess Considered for Commercialization by 1990

Volume III - Assessment of Advanced Indirect Coal Liquefaction Processes Considered for Commercialization by 1990

Volume IV - Review of Coal Liquefaction Technologies and Concepts Considered for Commercialization around 1995

Volume V - Process Developer Data and Information

This Volume III, includes an overview of advanced coal gasification processes selected by the Study Committee for prescreening and evaluation of indirect coal liquefaction of the designated Saskatchewan lignites (Stage I of the study). The prescreening and ranking of the coal gasification processes is presented in Volume I, "Summary and Process Comparison". An atmospheric, entrained-flow Koppers-Totzek gasifier was selected.

This Volume III also covers the techno-economic assessment of the following three indirect coal liquefaction alternatives using the selected Koppers-Totzek gasifier for the production of syngas:

STUDY REPORT ORGANIZATION (cont'd)

- pure methanol (99.9 wt% purity) production via low-pressure synthesis process developed by Imperial Chemical Industries (ICI process).
- production of gasoline via Mobil Methanol-to-Gasoline (Mobil-MTG) process using ICI process for methanol synthesis.
- production of gasoline and other hydrocarbons via SASOL-type Fischer-Tropsch process using Synthol fluidized-bed reactor.

The assessment is based on evaluation of a conceptual, integrated, self-sufficient commercial plant processing 25,000 tonnes per day of as-received Willow Bunch lignite for production of mainly transportation fuels such as gasoline, diesel and jet fuel. Typical proximate and ultimate analyses and ash distribution analysis for Willow Bunch lignite are presented in Appendix A.

No comparison and ranking between the three indirect coal liquefaction processing alternatives is presented in this Volume III as it is covered in Volume I.

The technical assessment for each of the processing alternatives includes:

- conceptual design of an overall plant configuration, overall material balance, product efficiencies, and utilities summary.
- product quality and comparative analysis.
- environmental considerations.

STUDY REPORT ORGANIZATION (cont'd)

The economic assessment includes:

- capital cost estimation and the total investment requirement for a 25,000-t/d plant located in Saskatchewan.
- estimate of annual operating, maintenance and production cost.
- product cost calculations using "Energy Value" and "Product Value" basis.

The basic sources used are process developers data:

- Information package prepared by Krupp-Koppers GmbH for SNC Inc., on gasification of Willow Bunch lignite using Koppers-Totzek gasifier configuration and low-pressure Imperial Chemical Industries (ICI) methanol synthesis, Krupp-Koppers Reference No. VK-1/Km-Mo-1813, June 1982. This report is attached in confidential Volume V, "Process Developers' Data and Information".
- Report prepared by Mobil Research and Development Corporation for the U.S. Department of Energy. "Research Guidance Studies to Assess Gasoline from Coal by Methanol-to-Gasoline and SASOL-type Fischer-Tropsch Technologies", DOE Report FE-2447-13, August 1978.

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1.0 OVERVIEW OF ADVANCED OR COMMERCIALLY PROVEN COAL GASIFICATION PROCESSES FOR PRODUCTION OF SYNTHESIS GAS

1.0 OVERVIEW OF ADVANCED OR COMMERCIALLY PROVEN COAL GASIFICATION PROCESSES FOR THE PRODUCTION OF SYNTHESIS GAS

"Indirect" coal liquefaction, as distinguished from "direct" coal liquefaction (hydrogenation) involves gasification of the coal to produce synthesis gas ($CO+H_2$) as the first step followed by conversion of the purified syngas to liquid hydrocarbons or to methanol in the subsequent steps. Therefore, first, an appropriate coal gasification process which is best suitable for production of synthesis gas from designated Saskatchewan lignites should be selected.

The six commercially proven or advanced coal gasification processes selected by the Study Committee for prescreening (Stage I of the Study) are listed in Table 1-1).

Table 1-1 shows that three major types of gasifiers (fixed bed, fluidized bed and entrained flow) used at present for production of syngas from coal are considered in the study. All gasification processes listed in Table 1-1 are considered in the oxygen-blown mode which is usually applied to production of syngas.

A single coal gasification process had to be selected for techno-economic evaluation of all the three indirect coal liquefaction alternatives considered in the study, i.e., coal-based Mobil MTG plant, SASOL-type Fischer-Tropsch plant and coal-based plant for production of pure methanol (Stage II of the Study).

The Lurgi, Koppers-Totzek and standard Winkler gasifiers have a history of reliable operation on a large commercial scale and are usually referred to in the literature as "first generation" gasifiers. The term "second generation" gasifiers is used to describe either the derivations from the original first generation design (e.g., BGC/L and HTW gasifiers) or new concepts (e.g., Texaco and Westinghouse gasifiers).

A brief overview of the coal gasification processes, selected for prescreening is given in the following pages.

Table 1-1 Advanced Coal Gasification Processes for Production of Synthesis Gas

Process	Gasifier Type	Process Developer
Lurgi (L)	Fixed Bed (dry bottom)	Lurgi Kohle and Mineralöltechnick, F.R.G.
British Gas Lurgi (BGC/L)	Fixed Bed (slagging)	British Gas Corporation U.K. and Lurgi, F.R.G.
Koppers-Totzek (GKT)	Entrained Flow	Krupp-Koppers, F.R.G.
Texaco (Texaco)	Entrained Flow	Texaco Development Corporation, U.S.A.
Winkler (HTW)	Fluidized Bed	Rheinishche Braunkohlenwerke AG (Rheinbraun), F.R.G.*
Westinghouse (W)	Fluidized Bed	Westinghouse Electric Corporation, U.S.A.

* Developer of high-pressure and high-temperature Winkler coal gasification process (HTW), a derivation from standard Winkler coal gasification process.

1.1 LURGI GASIFIER

The Lurgi gasifier is a first generation, pressurized, fixed-bed, dry-bottom (dry ash) gasifier. A typical diagram of the gasifier is given in Figure 1-1.

1.1.1 **Process Description**

The lump-size coal is fed to the gasifier via a lock hopper system at the top of the vessel and flows downwards countercurrent to ascending hot gases. As the coal descends, it flows successively through drying, devolatilization, gasification and combustion zones. The ash is withdrawn at the bottom of the gasifier by a slowly rotating grate and removed through locks at the base. Oxygen and steam are fed to the bottom of the coal bed through the rotating grate. The operating pressure in the gasifier is about 2.5 to 3.0 MPa. The temperature varies in the different zones, building up to a maximum (about 1000°C) in the bottom of the bed but limited to the ash melting point to avoid clogging the grate with clinker. Only a portion of the steam used to keep the temperature below the ash melting point in the bottom of the bed is converted. Therefore, there is a large amount of unconverted steam resulting in increased heat losses and expensive treatment of large volumes of condensed sour water. Raw gas exits from the gasifier at a relatively low temperature, typically in the range or 500°C to 650°C. Therefore, the gas contains hydrocarbons and condensible byproducts. The liquid byproducts (essentially tars, oils and phenols) amount to roughly 5 wt% of the coal feed.

1.1.2 Process Status

The Lurgi dry-bottom gasifier is a commercially proven process for generation of gas from coal. Over 100 gasifiers are in use in many large commercial installations. The 4-metre -diameter model is rated at 50,000 m^3/h of raw gas. A 5-metre model with 100,000 m^3/h capacity was installed at SASOL-I plant in mid-1980. The two largest installations made in recent years are at SASOL-II and SASOL-III plants in South Africa, each with 36 Lurgi gasifiers to produce synthesis gas for Fischer-Tropsch synthesis in Synthol reactors. High-ash African coal is used as the feedstock. Only about 70% of the coal at the plant gate is used for gasification due to the large amount of fines which are separated and sent to utilities. To solve the problem of gasification of fines, SASOL Ltd. is planning to install a 1200-t/d coal Westinghouse fluized-bed gasifier at SASOL-II plant (see details in Chapter 1.6, Westinghouse Coal Gasification Process).

The raw gas contains a large amount of methane since about 19% of the carbon in coal is converted to methane in the dry-bottom Lurgi gasifier. When operating at peak efficiency, the dry-ash Lurgi/Synthol combination yields over 50% of the output as Substitute Natural Gas (SNG). At SASOL-II (and at operational part of SASOL-III) these gases are reformed by partial oxidation in order to increase the liquid product yield. The overall thermal efficiency of the SASOL-II plant decreases from 55% - 58% to less than 40% due to recycling of SNG to the process for reforming.

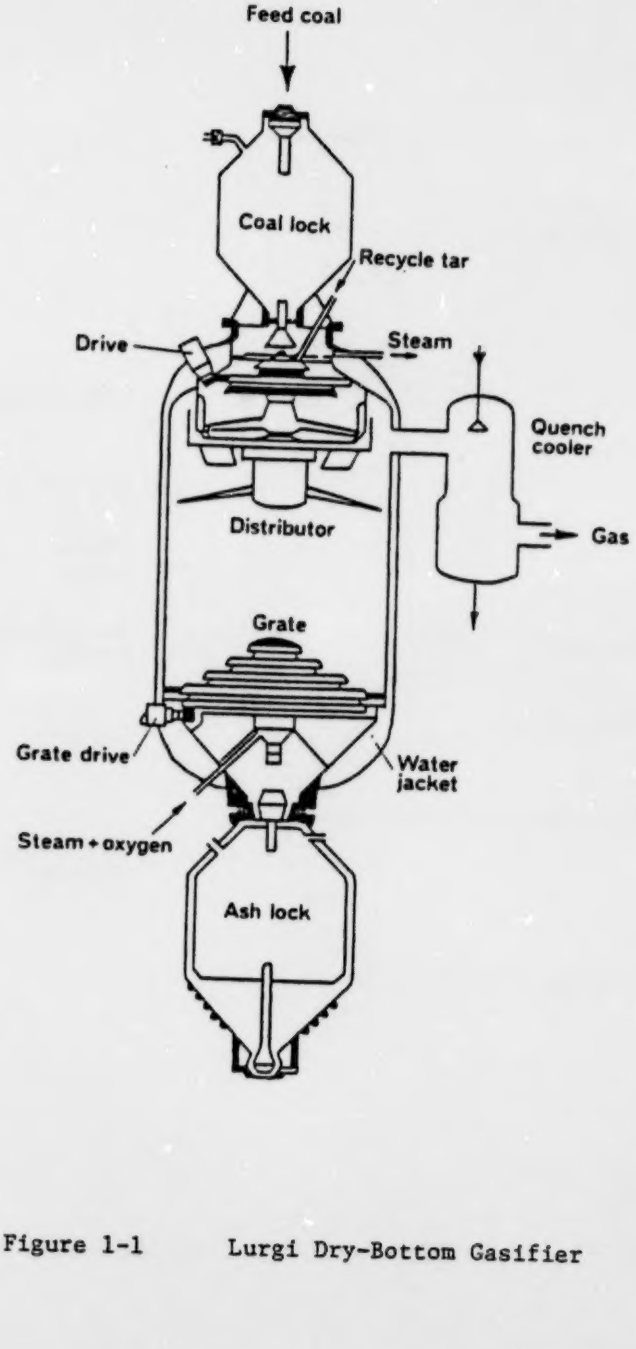

Figure 1-1 Lurgi Dry-Bottom Gasifier

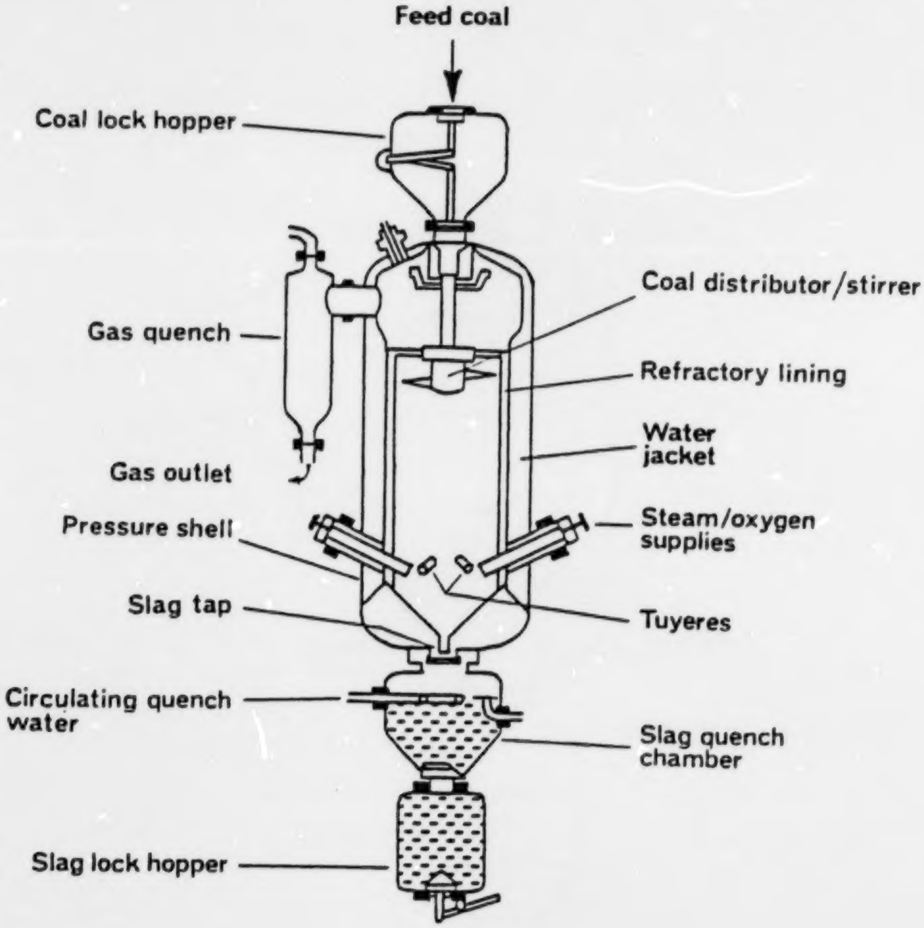

Figure 1-2 British Gas/Lurgi Slagging Gasifier

1.2 BRITISH GAS/LURGI GASIFIER

The British Gas/Lurgi gasifier is a derivation of the Lurgi dry-bottom gasifier. The main purpose for this development was gasificaton of the unreactive high-volatile bituminous coals, particularly those with low melting point of the ash and which showed poorer performance in the dry-bottom Lurgi gasifier.

1.2.1 **Process Description**

A simplified diagram of a British Gas/Lurgi gasifier is given in Figure 1-2. The British Gas/Lurgi gasifier is a fixed-bed, pressurized, slagging gasifier. The main features of this gasifier in contrast to the dry-bottom Lurgi are as follows:

- The bottom part of the gasifier is lined with refractory.
- The rotating grate is eliminated and the slag quench chamber is added.
- A blast-furnace type tuyères to inject steam and oxygen are installed.

The fuel bed is supported on the refractory hearth surrounded by a number of steam and oxygen tuyères with facilities for running off liquid slag through a tap hole into a chamber below, where it is quenched in water. The quenched slag is discharged from the pressure system through a lock hopper.

The change in mode of operation from dry ash to slagging has a significant effect on gasifier performance for high rank coals. The process steam consumption is reduced by up to 88% (i.e., only 12% to 15% of the steam is actually consumed, the rest is

1.2.1 **Process Description** (cont'd)

required to limit the temperature below the ash melting point in the Lurgi dry-bottom gasifier). This also allows the bed loading of the gasifier to increase before carry-over of dust from the gasifier becomes a problem. As a result, the gas output greatly increases (about three times) for a given size of gasifier. Thus, significant capital savings are possible, as well as improvement in the overall thermal efficiency of the process.

The gasifier operates at about 1500°C temperature and 3 MPa pressure. Increased amount of fines may be handled. Some of the fines, as well as tar and oils which are carried out from the gasifier with the gas can be recycled through the tuyères.

The syngas produced has lower hydrogen and higher carbon monoxide content than that from Lurgi gasifier, thus requiring "extra shift" to increase the hydrogen content before the syngas would be suitable for a synthesis reactor.

The performance of the British Gas/Lurgi slagging gasifier was proved during a 3-year test at Westfield with high-rank coals. Some of the data are presented in Table 1-2. A comparison between the slagging and non-slagging operation is also shown. However, this comparison may not be entirely representative since the coals were less suited to the non-slagging mode.

Table 1-2: Performance Data of British Gas/Lurgi Slagging Gasifier Compared with Non-Slagging Gasifier)

Gasifier	Slagging				Non-slagging
Coal Origin	Frances Scotland	Killoch Scotland	Ohio 9 U.S.A.	Pittsburgh 8 U.S.A.	Pittsburgh 8 U.S.A.
Size (mm)	25-6	25-6	25-6	30-6	30-6
Proximate Analysis (wt%)					
Moisture	7.6	7.7	5.4	3.2	4.6
Ash	5.2	4.5	21.3	7.5	7.7
Volatile Matter	32.5	33.8	32.5	35.1	37.4
Fixed Carbon	54.7	54.0	40.8	54.2	50.3
B.S. Swelling No.	1½	3½	6½	7½	7½
Caking Index (Gray-King)	B	E	G6	G8	G8
Operating Conditions					
Pressure, MPa	2.4	2.4	2.4	2.4	2.4
Steam/Oxygen Ratio (molar)	1.3	1.3	1.3	1.3	9.0
Outlet Gas Temperature, °C	482	468	410	515	675
Crude Gas Composition (vol%)					
CO_2	2.3	3.3	5.3	3.9	31.1
CO	57.5	56.4	53.7	55.5	18.0
CH_4	6.7	6.7	7.2	7.2	8.58
C_2H_6	0.4	0.5	0.6	0.1	0.7
C_2H_4	0.2	0.2	0.2	0.2	0.3
H_2	28.7	29.1	28.6	29.1	39.0
N_2	4.2	3.8	4.5	4.0	2.4
	100	100	100	100	100
GCV (MJ/m^3)	13.3	13.3	13.3	13.3	10.7
Derived Data					
Coal Gasification Rate (t/m^2h)	4.16	4.13	3.27	4.25	0.71
Steam Consumption (kg/kg MAF coal)	0.406	0.416	0.407	0.423	3.54
Oxygen Consumption (kg/kg MAF coal)	0.564	0.59	0.564	0.562	0.7
Crude Gas (MJ/kg MAF coal)	29.2	29.7	28.3	29.7	28.1
Tar and Oil (wt% MAF coal)	5.95	8.3	6.85	6.3	4.6
Net Gas Liquor (kg/kg MAF coal)	0.100	0.097	0.099	0.096	2.81

Source: G.F.I. Roberts, J. Hebden, C.T. Brooks and J.D. Sudbury, British Gas Corp., paper presented in the 14th World Gas Conference, May 1979.

1.2 BRITISH GAS/LURGI GASIFIER (cont'd)

1.2.2 **Process Status**

In 1975, a standard 2.8-m-diameter Lurgi dry-bottom gasifier was converted to a slagging mode of operation. For three years, this modified Lurgi gasifier, now called British Gas/Lurgi slagging gasifier, was tested on high-rank coals. British Gas Corporation is the licenser of the process. A 3-m-diameter gasifier which will gasify 600 to 800 t/d of coal is under construction at Westfield. The engineering design of two units, 1000 t/d coal each, to produce SNG at the Conoco demonstration plant in U.S.A. is completed.

British Gas Corporation is supporting Florida Power Corporation in a feasibility study for integration of a slagging gasifier (1000 t/d coal) with combustion turbines and exhaust heat recovery steam generators at the Higgins Plant in Pinellas County, Florida. British Gas Corporation is now prepared to grant licences for plants utilizing slagging gasifiers of sizes up to 3-m-diameter and will provide full commercial guarantees. In view of the results obtained at Westfield and the ongoing program, there is no technical objection to ordering plants at the present time based on 3-m-diameter gasifiers.

The British Gas/Lurgi slagging gasifier is at an advanced stage of development and can now be regarded as ready for commercial application. It is particularly suitable for high volatile bituminous coals and can find wide application in the areas of SNG manufacture, combined cycle power generation, medium Btu and synthesis gas production.

1.2.2 **Process Status** (cont'd)

Research and development of the British Gas/Lurgi gasifier was concentrated on the higher rank coals for which this process is most advantageous. No lignites have been evaluated at the British Gas Development Centre at Westfield.

As stated by the process developer*, British Gas Corporation, due to lack of operational experience on gasification of lignites, plant operations at Westfield would be necessary to obtain the design data, if Saskatchewan lignite is to be used as a feedstock.

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* Direct communication, November 1981.

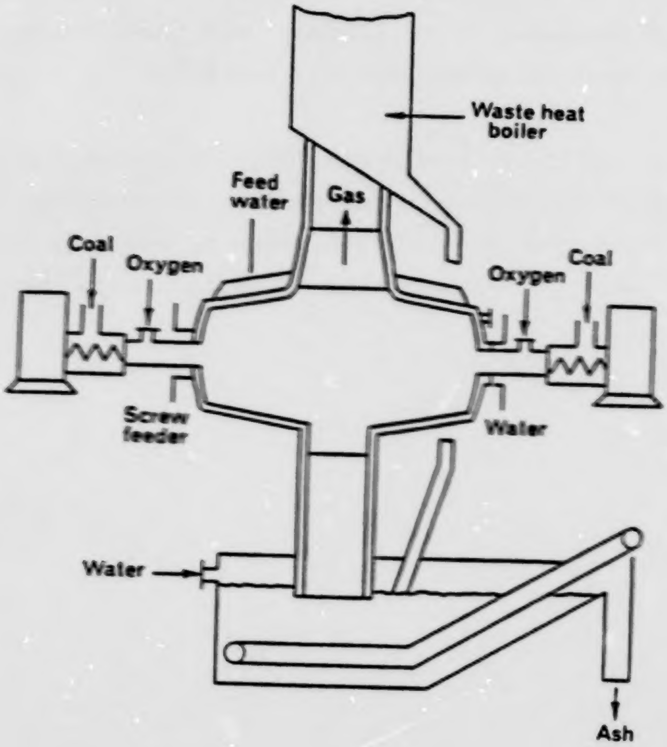

Figure 1-3 Koppers-Totzek (GKT) Gasifier

1.3 KOPPERS-TOTZEK GASIFIER

The Koppers-Totzek gasifier is a first generation, entrained-flow, gasifier. A simplified diagram of GKT gasifier is shown in Figure 1-3. The first commercial unit was built in 1952. At present, 79 commercial units are installed. Among them, six units are in a continuous commercial operation at the Ptolemais ammonia plant, Greece, converting high-ash lignite to 1,036,000 m^3/d synthesis gas.

1.3.1 Process Description

Predried and pulverized coal is continuously fed to the gasifier by screw conveyors together with steam and oxygen. The coal, steam and oxygen flow upwards concurrently, the coal particles being entrained in the gases as a dilute suspension. Coal drying, devolatilization, gasification and combustion occur very rapidly, virtually immediately, because of the high temperature in the reaction zone (approaching 1900°C). Very high carbon conversion is achieved (approaching 98% to 99%) and 40% to 60% of ash is discharged beneath the gasifier in the form of slag. The remainder is entrained as fly-ash with the gas produced and collected after the gas is cooled and washed out. Since coal is heated to the temperature above the ash fusion point, oxygen consumption in the entrained-bed gasifiers is higher than that for a fixed-bed or fluidized-bed gasifiers. On the other hand, due to the high operating temperature, the raw gas is practically free from methane, as well as from any condensible hydrocarbons (tars, oil) and phenols. The amount of methane in the raw gas is usually less than 0.1 vol.%. Therefore, the entrained-flow gasifiers are quite suitable for production of synthesis gas.

1.3.1 **Process Description** (cont'd)

Composition of the gas produced in Koppers-Totzek gasifiers processing lignite at the Ptolemais plant in Greece is presented below.

	vol%
CO_2	10.6
CO	60.0
H_2	26.9
N_2/Ar	1.9
CH_4	-
H_2S/COS	0.6

It may be seen that amounts of methane in the gas produced, as well as condensible hydrocarbons is nil. The H_2/CO ratio is about 0.5.

The operating pressure is about 0.05 MPa. Oxygen consumption is 0.3137 m^3/m^3 of gas produced.

1.3.2 **Process Status**

The Koppers-Totzek coal gasification process is commercially proven in many industrial complexes all over the world for the production of methanol, ammonia, hydrogen and fuel gas. The process was developed by Koppers-Totzek and, at present, the proprietor of the process is Gesellschaft fur Kohle Technologie (GKT) mbH, Essen, F.R.G. The licenser of this process is Krupp-Koppers GmbH, Essen, F.R.G. The process is known in the

1.3.2 **Process Status** (cont'd)

U.S.A. as the GKT coal gasification process. GKT gasification plants installed all over the world are listed in Table 1-3.

All types of coal can be processed in Koppers-Totzek gasifier and there are no limitations on fines in the feed coal. This includes high-ash-and-moisture content lignites such as Estevan (15.4 wt% ash MF coal) and Willow Bunch (21 wt% ash MF coal). The latter is proven by successful commercial operation of Koppers-Totzek gasifiers since 1964 at the Ptolemais plant, Greece, gasifying lignite containing 31 to 34 wt% ash (predried to 8% moisture). However, ash content in the coal (or char) fed to the gasifier should not exceed 40 wt% from economic and technological standpoints, as stated by process developer.

This gasifier is quite suitable for production of synthesis gas in an indirect coal liquefaction plant also because of absence of methane and condensible hydrocarbons and phenols in the raw gas produced.

Compared with pressurized entrained-flow gasifiers (e.g., Texaco), "atmospheric" Koppers-Totzek gasifiers have a lower specific gas output and also require additional energy for compression of raw gas to the elevated pressure at which subsequent units are usually operated (e.g., gas desulfurization, shift conversion units, etc.). However, the recent Krupp-Koppers study* indicates that for low-cost coals such as the designated

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* H. Staege , "Entrained Bed Gasification", Hydrocarbon Processing, March 1982.

Table 1-3 GKT Gasification Plants

	Gasifier Units Including Spare	Capacity CO + H_2 1000 m^3/d	Final Product	Year of Order
Mazingarbe, France	1	150	A.M.	1949
Ouiu, Finland **	5	280	A	1950/55
Nihon Suiso, Japan	3	210	A	1954
Puerites, Spain **	4	417	A	1954/61
Zandvoorde, Belgium	2	176	A	1955
Estarreja, Portugal	2	169	A	1956
Ptolemais, Greece **	6	1,036	A	1959/70
Suez, Egypt	3	778	A	1963
Mae Mon, Thailand	1	217	A	1963
Kutahya, Turkey	4	775	A	1966
Zeitz, DDR	2	360	A	1966
Kafue, Zambia **	4	877	A.M.	1967/75
Ramagundam, India	3	2,000	A	1969
Talcher, India	3	2,000	A	1970
Korba, India	3	2,000	A	1972
Modderfontein, South Africa	6	2,150	A.M	1972
San Jeronimo, Brazil	2	1,500	F	1979
Janina, Libiaz, Poland	3	2,700	F	1980
ICRC (SRC-1), Kentucky, USA	4	2,076	H	1980
TVA, Tennessee, USA	18	14,276	M.F	1981

A = Ammonia, M = Methanol, H = Hydrogen, F = Fuel gas
** including extensions

Source: H. Staege, Entrained-Bed Gasification, Hydrocarbon Processing, March 1982

1.3.2 **Process Status** (cont'd)

Saskatchewan lignites, the overall process economics may be favorable when using atmospheric entrained-flow gasifier for production of syngas.

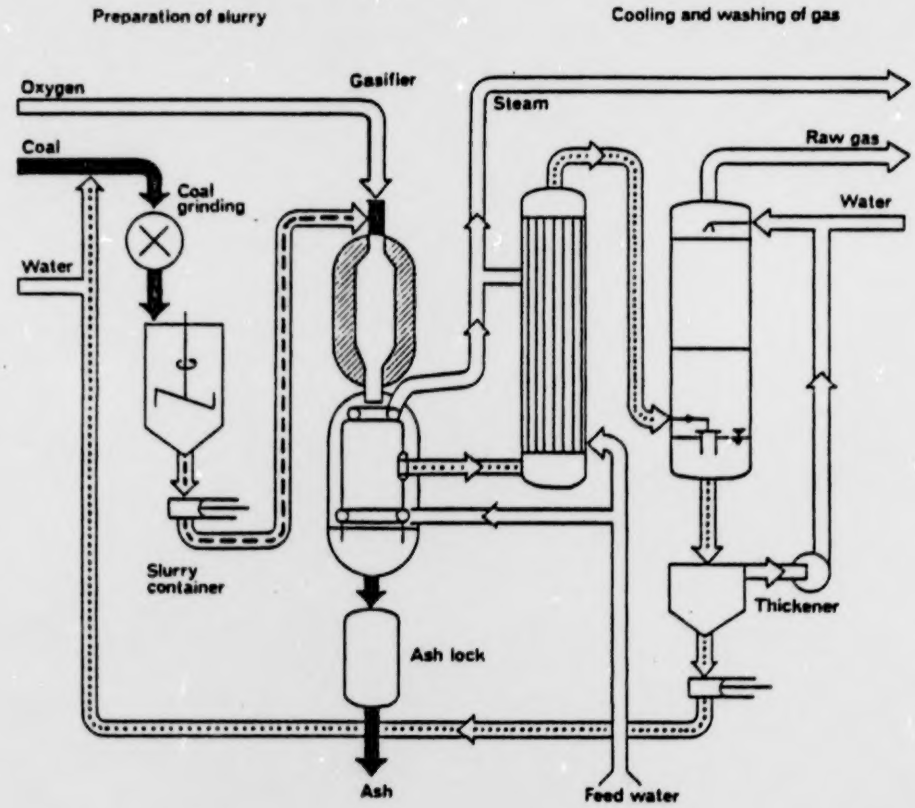

Figure 1-4 Texaco Coal Gasification Process (Waste Heat Recovery Mode)

1.4 TEXACO COAL GASIFICATION PROCESS

The Texaco coal gasification process is a second generation, pressurized, entrained-bed, slagging, downflow gasifier. A simplified diagram of Texaco gasifier is given in Figure 1-4. Whereas, most of the entrained-flow gasifiers (e.g., Koppers-Totzek, Shell) employ a dry coal feed system, the Texaco gasifier employs a wet coal feeding system using a concentrated slurry of ground coal in water.

1.4.1 **Process Description**

Figure 1-4 is a simplified diagram of the Texaco coal gasification process. A concentrated slurry of coal and water is fed into a refractory-lined pressure vessel through a special burner or injector where it is mixed with an oxidant (pure oxygen or air). The coal-to-oxygen ratio is adjusted so that the temperature within the gasifier is maintained at a level slightly higher than the fluid temperature of the coal ash. Under these conditions, the gasification reactions occur rapidly and 90% to 99% of the carbon contained in the coal is gasified in a few seconds. Hot gases exit from the gasifier along with molten slag and are cooled in one of the two alternate modes, the water quench mode or the waste heat recovery mode. The waste heat recovery mode shown in Figure 1-4 would typically be selected for process sequences in which methanol, or other petrochemicals are the desired end-products.

In this mode, the hot heterogeneous stream leaving the gasifier is directed downward through a radiant, water-jacketed cooler where high-pressure steam is generated as the gases are cooled. This cooler or waste heat boiler is designed to cool the gases and slag in the gas stream to a temperature well below the

1.4.1 **Process Description** (cont'd)

softening point of the coal ash. The majority of particulate matter in this stream is collected in the water-filled chamber at the bottom of the radiation section. Hot gases containing a small fraction of the particulate matter are then passed through a convection section where additional heat is recovered. At the exit of the convection section, the gases are scrubbed with hot water for final particulate removal and the resulting synthesis gas is ready for further treatment.

The slag from the coal remains in the water where it settles rapidly and is withdrawn periodically through a water-sealed lock hopper system. Unconverted carbon and some fine particles are present in the circulating water and are removed from the water by passing the water stream through a settler. Material recovered in the settler may be discarded or, optionally, recycled back to prepare additional slurry feed for the gasifier. The material removed through the lock hopper is typically passed over a screen where the coarse particulate containing normally about 99% ash are removed from the system and discarded. Extensive tests have been made on this "coarse slag" stream and have shown that disposal by landfill poses minimal environmental impact. The underflow from the screen or "fine slag" usually contains some unconverted carbon and, optionally, can be fed back to the slurry preparation system.

Water recovered from all the streams is recycled to the process or to slurry preparation and the only water discharge from the process is that required to maintain an acceptable level of total dissolved solids in the circulating water. The build-up of solids in the circulating water stream is dependent upon the

1.4.1 **Process Description** (cont'd)

composition of the nonhydrocarbon material in the coal feed. Depending on the content of water soluble material in the feed coal, a water purge or blowdown stream ranging from 0.25 to 1.0 lb per pound of coal fed is typical.

All materials leaving the gasifier are exposed to relatively high temperatures in the order of 1200 to 1500°C. At this temperature level, partially gasified or partially oxidized carbonaceous material is extremely unstable. Therefore, the undesirable components, such as phenols,cresols and other partially oxidized or carcinogenic hydrocarbons are not found. In addition, the ability to recycle the small fraction of unconverted carbon back to the slurry feed greatly reduces the problem of byproduct disposal. The high temperature and high pressure within the gasifier lead to rapid gasification reactions and therefore the gasifier itself is capable of relatively high throughputs per unit volume.

1.4.2 **Process Status**

In the last few years, the Texaco coal gasification process has won wide acceptance as one of the few processes which is ready for commercialization today. The Texaco technology has been developed over a period of more than thirty years. The process development work was initiated at the Texaco Montebello Research Laboratory near Los Angeles after World War II and in 1948 a pilot unit was constructed at the laboratory. In the early development years, the Texaco synthesis gas generation Process in which syngas is manufactured from heavy residual oil and the companion Texaco coal gasification process were developed side by

1.4.2 **Process Status** (cont'd)

side. Work on residual oil gasification led to the installation and start-up of the first commercial residual-oil feed Texaco synthesis gas generation unit in 1956, and work on coal gasification resulted in construction of a prototype unit in Morgantown, West Virginia. The coal gasification plant at Morgantown was operated during 1956 and 1957, at the end of which time it was determined that the process could not compete economically with low-cost natural gas. In the meantime, residual oil gasification projects became quite attractive and the technology was licensed throughout the world, resulting in construction of more than 80 commercial plants.

During the period following the demonstration work at Morgantown, development efforts at the Montebello Research Laboratory were directed mainly toward oil gasification and it was not until the oil embargo in 1973 that the emphasis was redirected and efforts were devoted to coal gasification.

In 1973, the laboratory converted one of the oil gasification pilot plants to a coal gasification unit and process development work on coal was re-established. Facilities at the Montebello Research Laboratory include two 15 to 20-t/d coal gasifiers operating at pressures ranging from 2.3 to 8.2 MPa.

Test runs in the pilot units have been made on a wide variety of coals and on residual products from several of the coal liquefaction processes, as well as coke from refining conventional petroleum residue. North Dakota, Texas and Ptolemais lignites have been tested at Montebello pilot plant, in particular.

1.4.2 **Process Status** (cont'd)

Slurries containing more than 60% coal (MF basis) have been handled. The use of certain surface-active agents can reduce the slurry viscosity significantly and, in some cases, permit concentrations of up to 70% dry solids. The Texaco Development Corporation reported to C.F. Braun Co.* that 55% concentration of coal in slurry was achieved when using pretreated North Dakota lignite.

The accelerated development of Texaco Coal Gasification following the oil embargo in 1973 led to another demonstration plant in Germany at the Ruhrchemie Petrochemical Complex near Oberhausen. This plant, jointly designed by Texaco, Ruhrchemie, and Ruhrkohle, is rated at 165 t/d of coal. Since the start-up of this plant in January 1978, more than 40,000 tons of various bituminous coals have been gasified in 7500 hours of operation. The designed operating pressure is about 3.6 MPa.

Another demonstration plant using an oxygen-blown Texaco gasifier is the TVA plant (Tennessee Valley Authority's Ammonia-from-Coal Plant), in Muscle Shoals, Alabama. The TVA plant, which is designed to gasify bituminous Illinois No. 6 coal at a rate of 190 t/d and at pressure of about 4.0 MPa, is in the testing phase. Mechanical problems related to the automatically operated water-sealed lock hopper system and valves used for ash discharging were experienced. However, the valves installed at the Ruhrchemie/Ruhrkohle plant have cycled more than 30,000 times to date with only minor maintenance problems.

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* Production of Methanol from Lignite, EPRI Report AF-1161, Part II, 1979.

1.4.2 **Process Status** (cont'd)

Data from the Montebello pilot plant and Oberhausen demonstration plant are being used to design and construct a number of larger plants incorporating the Texaco coal gasification process. The two most important projects underway at present are the Cool Water Coal Gasification complex, planned for construction in the Southern California and the Tennessee Eastman project.

The Cool Water Coal Gasification program is organized to design, construct, start up and test a modular commercial coal gasification combined-cycle plant for a period of 7 years. The project, estimated to cost $300 million is designed to gasify 1,000 t/d of a western bituminous coal in an integrated combined-cycle generation facility to produce 100 MW of power for distribution in Southern California.

The Tennessee Eastman plant which is also in the final engineering stage will be the first full-scale commercial operation of the Texaco coal gasification process. Syngas from the plant will be used for a wide variety of chemical products most of which are related to the line of photographic materials Eastman manufactures.

A number of lignites including North Dakota, Texas, Ptolemais (Greece) have been tested in the Texaco coal gasification pilot plant (15 to 20 t/d of coal) at Montebello. Positive results have been reported by the process developer. Thus, the feasibility of gasifying lignites in the Texaco gasification process has been established.

1.4.2 **Process Status** (cont'd)

Preliminary assessment on suitability of Saskatchewan lignites carried out by the Texaco Development Corporation, at the request of SNC and based on proximate and ultimate analyses of these lignites and ash analyses, predicted no problems in application of the Texaco coal gasification process. In particular, simulated performance data predicted concentration of as-received Willow Bunch lignite in the slurry at about 48% to 49% (MF basis) without predrying of coal.

Predicted composition of raw gas is shown below:

	vol %
CO	16.47
H_2	19.69
CO_2	19.58
H_2O	43.69
CH_4	0.09
Ar	0.10
N_2	0.15
H_2S	0.22
COS	0.01

The data shows that the product gas contains negligible methane. The H_2/CO ratio is 1.19. Both factors are positive for production of synthesis gas to be used in a indirect coal liquefaction plant.

1.4.2 **Process Status** (cont'd)

However, a test on grinding and slurrying of a representative sample of Willow Bunch lignite in water carried out later by Texaco has shown that in order to achieve the predicted solid concentration, coal has to be completely dried before slurrying (vacuum drying was used in the test to achieve 49% concentration of coal in the water slurry). Taking into account the high moisture content of Saskatchewan lignites (35 wt%), the extent of coal drying required before slurrying would result in substantial additional expenses (about $7 to $9 per tonne of coal) and also may cause decrease in coal reactivity due to the possible damage of porous structure during coal drying. The process of such a deep drying of coal from 35 wt% moisture is not proven commercially. Therefore, at present, suitability of Saskatchewan lignites for gasification in the Texaco coal gasification process is questionable.

It should be noted, however, that further investigation has been planned by the Texaco Development Corporation on pretreatment of lignites, in particular the use of certain additives to increase the solid concentration of coal-water slurry. These improvements may reveal that the process is attractive for gasification of high moisture and ash lignites.

1.5 H.T. WINKLER GASIFIER

The high-temperature, high-pressure Winkler process (H.T. Winkler) is a second generation gasifier and a derivation of a commercially-proven "standard" fluidized-bed Winkler gasifier which operated at a low pressure (about 0.12 MPa) and relatively low temperature (800 to 900°C). Commercial standard Winkler gasifiers were installed in many countries all over the world, in Spain, Yugoslavia, Japan, Turkey, India and Germany. From 1956 to 1964, two standard Winkler gasifiers, each producing 17,000 m^3/h of raw gas were operated at Union Rheinische Braunkohlen Kraftstoff, a subsidiary of Rheinische Braunkohlenwerke AG, at an operating pressure of about 0.15 MPa and temperature of 800 to 900°C.

In the Winkler gasifier, ground coal is suspended in a fluidized condition by oxygen and steam which flow upwards. Coal undergoes drying, devolatilization, gasification and combustion at essentially constant temperature because of rapid mixing characteristics of the fluidized bed. Due to a more moderate operating temperature than in entrained-flow gasifiers, the residence time is greater and oxygen consumption is lower (usually the operating temperature in fluidized-bed gasifiers is below ash melting temperature). Coal gasification in fluidized-bed conditions offers good temperature control and high heat transfer rates.

Low-rank coals with high percentage of fines and high-ash content (up to 50 wt%) were successfully gasified in a commercial standard Winkler gasifier. Predrying of coal is required only to the degree necessary to ensure adequate free-flowing coal characteristics and economic oxygen utilization (for certain moisture levels, drying of coal in the gasifier with oxygen is more expensive than an external drying system). Typically, coal is predried to between 12 and 16 wt% moisture. Some quantities of methane (but less than in fixed-bed gasifiers) is contained in the gas produced but higher hydrocarbons are generally neglible. However, due to the relatively low temperature at which the standard Winkler gasifier is operated, hydrogen and carbon monoxide content as well as carbon conversion are relatively low. The low operating pressure also contributes to lower capacity per gasifier and higher energy requirement for compression of gas produced. Due to these disadvantages, no sales of standard Winkler gasifiers are known since 1960. To overcome these disadvantages, Rheinbraun is developing a high-temperature, high-pressure gasifier (H.T. Winkler process) specifically for generation of synthesis gas for methanol production from low-rank coals. Most of the development work is based on processing of Rhenish brown coal of low ash (3 wt%) and high moisture content (60 wt%). Before feeding to the gasifier, coal is predried to about 12 wt% using the proprietary Rheinbraun technology.

1.5.1 Process Description

A simplified diagram of the H.T. Winkler process is shown in Figure 1-5. In the HTW process, coal ground and predried to 12 wt% is fed into the gasifier via a feed lock hopper system and a screw conveyor. In the fluidized bed, coal reacts with oxygen and steam which are injected on several levels. The product gas discharged from the gasifier carries entrained particles. Larger particles are separated from the gas in the first separator and recycled to the gasifier. Fine particles (mostly fly ash) are separated in the second separator (not shown in Figure 1-5) and removed via water-cooled screw conveyor and lock hopper. Ash is discharged via ash water-cooled conveyor and lock hopper. The gas leaving the gasifier is cooled in a waste heat boiler.

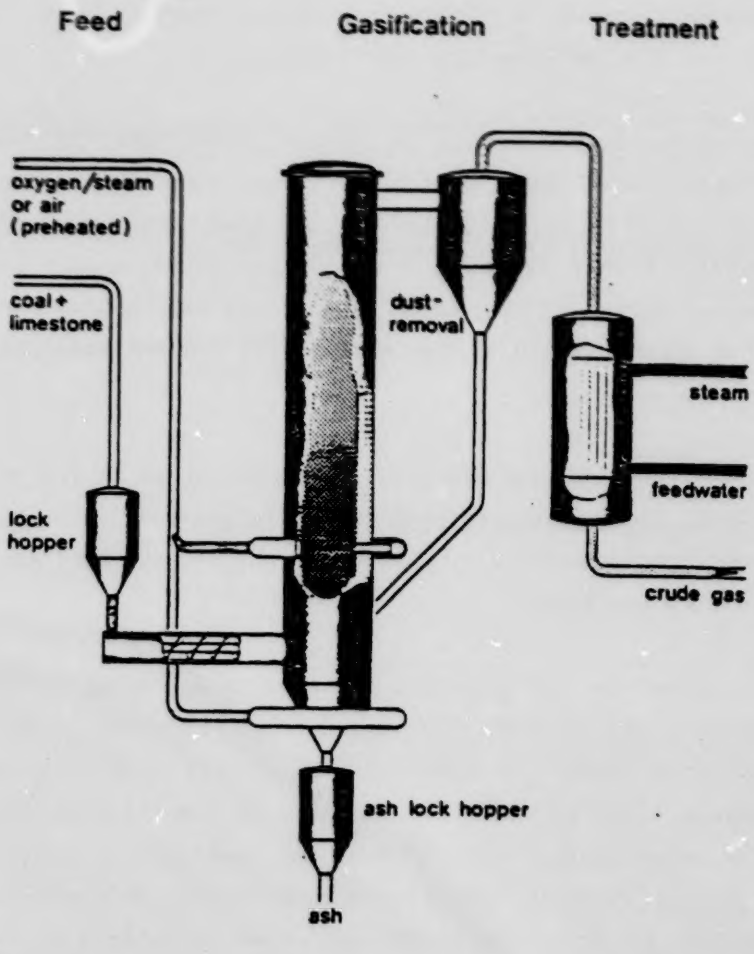

Figure 1-5 High-Temperature Winkler Process

1.5.2 Process Status

The development of the H.T. Winkler process is based on the experience gained by Rheinbraun through operation of two commercial standard Winkler gasifiers processing Rhenish brown coal from 1956 to 1964 at Wesseling near Cologne.

The first tests leading to further development of the Winkler process were carried out at the Rheinische Westfalische Technische Hochschule, Aachen, in 1965 using the bench-scale gasifier with a capacity up to 50 kg/h of MF coal. Since 1978, the experimental program is also being carried out at the HTW pilot plant located at the site of the Frechen briquetting plant near Cologne.

The capacity of the HTW pilot plant is up to 32 t/d of coal (MF basis). Operating pressure and temperature is around 1 MPa and 1100°C respectively. Steam and oxygen or air are used as gasification agent.

The plant is fed with dried brown coal having about 18 wt% moisture as required for briquetting purposes. In a spiral pneumatic drier, the water content of coal is further reduced to between 8 and 12 wt%. Test results of the pilot plant operation with steam/oxygen mixtures used as gasification agent compared with the optimum values obtained in the atmospheric Winkler gasifier of Union Kraftstoff are shown in Table 1-4. It may be seen that carbon conversion was increased to over 95% and the specific synthesis gas output of the gasifier was more than doubled.

Table 1-4 Operating Conditions and Parameters for Standard and H.T. Winkler Gasifier.

Operating Condition and Parameters	Standard Winkler*	H.T. Winkler
Temperature, °C	950	990
Pressure, MPa	0.12	1.0
Oxygen, m^3/kg MF coal	0.398	0.38
Steam, m^3/kg MF coal	0.167	0.41
Raw Gas Yield, m^3/t MF coal	1396	1483
Specific Output of Gasifier, m^3/m^2h	2122	5004
Carbon Conversion	91	95

Sources: H. Teggers et al, "Actual Status of the Rheinbraun High-Temperature Winkler (HTW), International Coal Conversion Conference", August 1982, South Africa

* Flesch/Velling, "Die Vergosung von Kohle in Winkler Generator Erdal und Kohle", Sept. 1962.

1.5.2 **Process Status** (cont'd)

By late March 1982, nearly 5000 tonnes of brown coal (MF basis) had been gasified in the HTW pilot plant. The decision was made to build an HTW demonstration plant at the site of the Rheinbraun Ville/Berreurath briquetting plant for converting the local Rhenish coal into synthesis gas. The plant is planned to be built in two stages:

- construction of the first processing line with one HTW gasifier having the capacity to process 55 t/h MF coal and produce 37,000 m^3/h of raw gas. This first line is scheduled for start-up in 1984.

- extension of three more processing lines with three HTW gasifiers. Total capacity (including coal to be used for the energy supply) is 280 t/h MF coal and raw gas output is 147,00 m^3/h. Start-up is scheduled in 1988.

1.6 WESTINGHOUSE FLUIDIZED-BED GASIFIER

Since 1972, the Westinghouse Electric Corporation, under sponsorship of the U.S. Department of Energy and its predecessor agencies, and more recently, in conjunction with the Gas Research Institute, has been developing a fluidized-bed coal gasification process. Most of the research and development work has been carried out at the Westinghouse Research Centre in Pittsburgh and at a 24-t/day pilot plant at Waltz Mill, Pennsylvania. Feasibility studies for several commercial size plants using the Westinghouse gasification process are underway (e.g., Keystone project, NASA Lewis Research Centre project, Westinghouse Fairmount Lamp project). However, until the recent time, the commercialization prospects for production of syngas from coal using Westinghouse process were quite uncertain.

At the end of 1981, the Westinghouse Electric Corporation took a "fast-track" approach in the development of the process. It was announced that Westinghouse is developing a commercial 1200-t/d coal gasification unit to be installed at the SASOL-II complex in South Africa with the start-up scheduled in early 1985. Because of this development, it may be expected that the Westinghouse process would reach readiness before 1990.

1.6.1 **Process Description**

A simplified diagram of the Westinghouse gasifier is given in Figure 1-6. The Westinghouse coal gasifier consists of a single stage, pressurized, fluidized-bed reactor. The gasifier is operated at about 950°C temperature and 3 MPa pressure. The as-received coal is ground and dried to a free-flowing condition (about 15 wt% for lignite). Coal is pneumatically transported from lock hoppers into the gasifier through a central feed tube

1.6.1 **Process Description** (cont'd)

where it is combusted and combined with a stream of oxygen or air and steam. The combustion of a portion of the incoming coal provides the necessary heat to devolatilize the remainder of the coal to form char and to combine the char with steam to form hydrogen and carbon monoxide. As the bed of char circulates through the jet formed at the end of the central feed tube, the carbon in the char is consumed by combustion and gasification, leaving particles that are rich in ash. The fluid dynamic design of the gasifier allows the ash-rich particles to agglomerate and form large, dense masses which defluidize from the bed and collect in the ash annulus. These agglomerates are cooled and removed through a lock-hopper system.

Raw product gas from the gasifier passes through cyclones to recover entrained particulates which are recycled to the gasifier. The solids-free gas passes through heat exchangers and is then quenched in a water venturi scrubber.

1.6.2 **Process Status**

As it was mentioned earlier, Westinghouse Electric Corporation has been developing its coal gasification process since 1972. Since 1975, a 24-t/d pilot plant at Waltz Mill, Pennsylvania is in operation. The pilot plant has logged over 8,000 hours of hot operation and processes a wide variety of coal feedstocks, including Texas lignite. According to the process developer's information obtained through direct communications, operating experience in the pilot plant has proved that run-of-mine coal

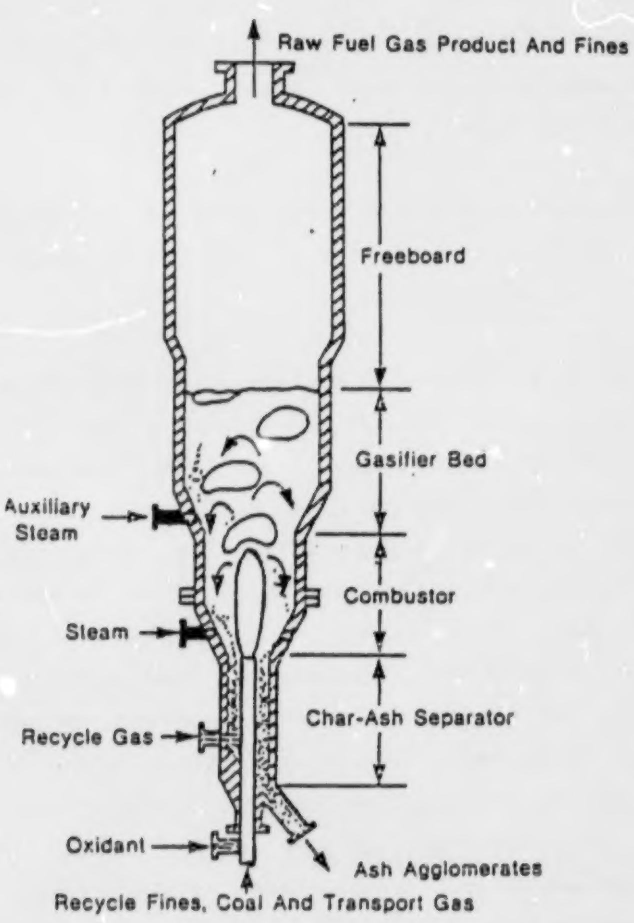

Figure 1-6 Westinghouse Gasifier

1.6.2 **Process Status** (cont'd)

containing sufficient moisture can be processed. Also coals with a wide range of sulfur and ash content can be used. The design of the system allows the Westinghouse fluidized-bed reactor to process both caking and non-caking coals, reactive and relatively non-reactive coals, and coals with high moisture content approaching 24 wt%.

Westinghouse Electrical Corporation is now developing the 1200-t/d gasification demonstration unit to be installed at SASOL-II complex in South Africa.

At the International Coal Conversion Conference held in South Africa in August 1982, it was announced that the project will be developed in five phases. First, Fluor Engineers and Constructors are doing the detailed design work for the project and should finish it in late 1983. The second construction phase will start in July 1983 and be completed in November 1984. The third phase of development will include a three-month start-up period followed by a demonstration period extending from early 1985 to mid-1986. The fourth phase will be an "optimization" period during which Westinghouse will make any necessary changes and then run the gasifier for another year. Finally, SASOL will take over operation of the unit. SASOL would use this ash-agglomerating, pressurized fluidized-bed gasifier to demonstrate the gasification of their coal fines, since Lurgi gasifiers which are in operation at the three SASOL plants have limitations on processing of coal fines.

1.6.2 **Process Status** (cont'd)

The demonstration unit will operate in a slip-stream mode using existing utility facilities. The demonstration unit will include coal preparation-feed, ash-removal, fines collection-recycle, heat-recovery and gas-quench systems.

Based on the process developer's positive experience in testing of lignites in the Westinghouse coal gasification process, it may be assumed that the Saskatchewan lignites are, in principle, suitable feedstocks for the Westinghouse pressurized, fluidized-bed gasifier.

The process developer also confirmed the general suitability of Saskatchewan lignites. However, to determine the quantitative process data, Westinghouse Electric Corporation requested to carry out a preliminary coal test which is beyond the scope and budget of the current study.

Table 1-5 Indirect Coal Liquefaction Cases Evaluated by Foster Wheeler

No.	Case	Gasifier	Synthesis	Additional Processing
1	I	Lurgi	M-MTG	-
2	IA	Lurgi	M-MTG	SNG Reforming
3	II	Lurgi	F-T	-
4	IIA	Lurgi	F-T	SNG Reforming
5	III	Winkler*	M-MTG	-
6	IV	Winkler*	F-T	-
7	V	Koppers-Totzek	M-MTG	-
8	VI	Texaco	M-MTG	-
9	VII	British Gas/ Lurgi	M-MTG	-
10	VIIA	British Gas/ Lurgi	M-MTG	SNG Reforming
11	VIII	British Gas/ Lurgi	F-T	-
12	VIIIA	British Gas/ Lurgi	F-T	SNG Reforming

* Standard Winkler gasifier was evaluated instead of HTW because the required information on HTW was not available.

1.7 PROCESS COMPARISON AND RANKING

Ranking of the above-described six coal gasification processes for indirect liquefaction of designated Saskatchewan lignites is presented in Volume I, Chapter 4. The ranking is based on technical assessment of the processes and suitability of designated Saskatchewan lignites for production of synthesis gas in the processes considered.

GKT coal gasification process including entrained-flow Koppers-Totzek gasifier was selected as a single gasification process for techno-economic assessments of all the indirect coal liquefaction alternatives considered in the study.

In addition, a preliminary evaluation of indirect coal liquefaction via Mobil-MTG (M-MTG) and SASOL-type Fischer-Tropsch (F-T) plants in combination with some of the gasifiers listed in Table 1-5 was carried out by Foster Wheeler.* The details and results are presented in Annex A.

Estevan lignite was assumed as a gasifier feedstock. British Engineering units were used in this evaluation. Preliminary economic information is presented in mid-1982 U.S. dollars. Twelve cases considered are listed in Table 1-5.

Summary of this preliminary evaluation is presented below.

According to the product criteria described earlier, maximum amount of gasoline and other liquid fuels and minimum (if any) gaseous hydrocarbons has to be produced via indirect coal liquefaction.

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* This work was done by Foster Wheeler Energy Corporation, N.J., F.W. report to SNC, November 1981, Job. No. B28-9011.

Under these product criteria, application of dry-bottom Lurgi and British Gas/Lurgi gasifiers for both Mobil-MTG and Fischer-Tropsch processes (Cases I, II and VII, VIII respectively) should be eliminated due to high amount of SNG and correspondingly low amount of gasoline in the product yield. For example, in Lurgi/M-MTG combination, the estimated SNG content in the product yield as a percentage of overall efficiency is 40.76%. In a Lurgi/F-T combination, the estimated SNG content is 53.44% (see Annex A).

Maximization of liquid product yield using Lurgi or British Gas/Lurgi gasifiers with reforming of SNG produced (Cases IA, IIA, VIIA and VIIIA) would result in sufficient decrease in overall efficiency. For example, for a Lurgi/M-MTG combination, the estimated overall thermal efficiency excluding input electricity is 41.0% with SNG reforming versus 52.8% without SNG reforming. Similarly, for a Lurgi/F-T combination, the corresponding decrease in efficiency is from 48.7% to 33.8%.

Application of Texaco or Koppers-Totzek gasifiers in combination with the F-T process would result in lower amount of SNG produced since there is practically no methane in the raw gas generated in entrained-flow gasifiers. The amount of SNG would be only that generated in F-T process itself. The Winkler/F-T combination still would result in a sufficient amount of SNG in the product yield.

In combination with the Mobil-MTG process, about 93% of gasoline is estimated in the product yield (as a percentage of overall efficiency) for application of Koppers-Totzek or Texaco entrained-flow gasifiers or Winkler fluidized-bed gasifier (Cases V, VI and IV).

Examination of the results of this preliminary evaluation leads to the conclusion that entrained-flow, pressurized gasifier should have the first priority for gasification of Saskatchewan lignites. Unfortunately, as discussed earlier (see Subsection 1.4.2), application of the entrained-flow pressurized Texaco coal gasification process seems to be uneconomic for Saskatchewan lignites due to the type of coal feeding system used in this process. Under these circumstances, the GKT coal gasification process including atmospheric, entrained-flow K-T gasifier appears to have the highest priority for gasification of designated Saskatchewan lignites as a single gasification process for all the indirect coal liquefaction alternatives considered in the study. These conclusions coincide with the prescreening and ranking of coal gasification processes presented in Volume I, Chapter 4.

The techno-economic assessments of three indirect coal liquefaction alternatives (Stage II of the Study) using GKT coal gasification process, as a first step in production of syngas, are presented as follows:

Chapter 2 - Technical Assessment of a Coal-Based Plant for Production of Pure Methanol

Chapter 3 - Technical Assessment of a Coal-Based Methanol-to-Gasoline Plant via Mobil-MTG Process

Chapter 4 - Technical Assessment of a Coal-Based, Sasol-Type Fischer-Tropsch Synthesis Plant

Chapter 5 - Environmental Aspects of the Indirect Coal Liquefaction Alternatives

Chapter 6 - Economic Assessment of the Indirect Coal Liquefaction Alternatives.

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2.0 TECHNICAL ASSESSMENT OF A COAL-BASED PLANT FOR THE PRODUCTION OF PURE METHANOL

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2.0 TECHNICAL ASSESSMENT OF A COAL-BASED PLANT FOR THE PRODUCTION OF PURE METHANOL*

2.1 OVERALL CONFIGURATION OF A CONCEPTUAL, SELF-SUFFICIENT, COAL-BASED METHANOL PLANT

The overall configuration of the commercial coal-to-methanol plant includes coal preparation, coal gasification by Koppers-Totzek entrained-flow gasifier and methanol synthesis by ICI process followed by raw methanol distillation along with the required supporting systems, utilities supply and offsite facilities to comprise an integrated, self-sufficient operation.

The plant is a complete grass-root facility requiring coal and water as basic raw materials. Catalysts and chemicals are also needed but in relatively small quantities. Part of the steam required is generated by recovering the process heat. The balance is generated in offsite boilers by burning coal. Electricity is generated within the plant by steam-driven turbines.

A list of processing areas or sections of a conceptual, self-sufficient coal-to-methanol plant is presented in Table 2-1 and the interrelationship between these processing areas is shown in the overall block flow diagram in Drawing 2-1. An overall material balance for main process streams around each processing area or section is also included in the Drawing 2-1.

An outline of processing areas of the overall block flow diagram along with a summary of product yield, thermal efficiency and utility balances is given in the following pages. A detailed description of Koppers-Totzek gasification of Willow Bunch lignite to produce synthesis gas and methanol along with process flow diagrams prepared by Krupp-Koppers GmbH is presented in a confidential Volume V, "Process Developers Data and Information".

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* to meet gasoline blend stock specifications (99.95 wt% purity)

2.1.1 Coal Preparation (Section 100)

The coal preparation section includes coal handling and storage facilities along with coal grinding, classifying and drying equipment to provide coal of desired specification for the following processing areas:

- Koppers-Totzek gasifiers (Section 200) for synthesis gas production; 70% to 90% minus 0.075 mm size, and 8 wt% moisture content.

- Coal-fired boilers for steam and power generation (Section 900); no size reduction but drying up to equilibrium moisture content of 28.5 wt% using low grade energy (e.g., air drying) that would otherwise be rejected.

The as-received Willow Bunch lignite of approximately minus 2-cm size and containing about 35% moisture is supplied to the plant battery limits by unit rail cars. In the coal receiving station, the rail cars dump the coal into hoppers located below the track. Coal is withdrawn from the hoppers by means of vibrating feeders and discharged onto a system of belt conveyors. The conveyor system includes facilities for coal sampling, weight-recording and removal of tramp iron by magnetic separators.

Table 2-1 Processing Areas of a Conceptual, Commercial Coal-To-Methanol Plant

SECTION NO.	DESCRIPTION
100	COAL PREPARATION Storage and Reclaiming Crushing and Drying
200	COAL GASIFICATION Lignite Distribution and Gasification Waste Heat Recovery Raw Gas Washing Slag Removal
300	GAS TREATMENT Raw Gas Compression Rectisol Sulfur Removal Shift Conversion Rectisol CO_2 Removal
400	METHANOL SYNTHESIS
500	OXYGEN PLANT
600	EMISSION CONTROL Claus Sulfur Recovery Plant SCOT Tail-Gas Processing Unit
700	EFFLUENT CONTROL
800	PRODUCT BLENDING, STORAGE AND SHIPPING
900	UTILITIES Raw Water Treatment and Distribution Cooling Tower and Cooling Water System Boiler Feedwater Preparation Condensate Treatment Steam Generation and Distribution Fuel Gas Purification and Distribution Plant-Air, Instrument Air and Inert Gas System Flare System
1000	OFFSITES
1100	PRODUCT UPGRADING AND/OR REFINING Distillation

2.1.1 Coal Preparation (Section 100) (cont'd)

Under normal conditions, the coal is conveyed to two coal storage silos, one located within the coal preparation section and the other near the offsite facilities. Surplus incoming coal is conveyed to an adjacent dead-storage area where two stock piles, each of approximately 400,000 tonnes capacity, are prepared to meet 30-day coal requirements. During emergency or shortfall operation, coal may be reclaimed by a travelling-wheel stacker-reclaimer of 1200-t/h capacity and sent to the coal silos by the reversible conveyor system. In the possible event of a problem with a reclaimer, coal may be retrieved by means of mobile equipment through an emergency reclaiming hopper.

The raw lignite from the storage silo is first transferred to lignite bunkers in the coal preparation unit through a pneumatic conveyor system. The lignite preparation unit has six parallel lines each equipped with one tube mill and can be operated independently. Lignite is ground in the tube mill where it is simultaneously dried to 8 wt% moisture with co-current flow of hot drying gases. The hot gases are produced by burning a portion of ground lignite in a direct combustion heater. The temperature of the hot gas is maintained around 450°C to keep coal particulate temperature at 85°C to prevent devolatilization. Dried lignite is passed through a sifter to separate oversize material which is fed back to the tube mill. The undersize lignite is separated from hot gases in the cyclones and sent to the lignite feed bunkers. Separated gases from the cyclones are recycled through an electrostatic precipitator where the dust content of the gas is reduced to 150 mg/m^3.

2.1.2 Coal Gasification (Section 200)

Dried and pulverized lignite is pneumatically conveyed from the storage bunkers (Section 100) to the gasifier service bins (two at each 4-headed gasifier) each of which in turn feeds a pair of feed bins. From each feed bin, the lignite is metered by two feed screws into a mixing zone where it is entrained with oxygen and fed into the gasification burners. Conveying nitrogen-containing lignite dust from gasifier service bins is sent to the electrostatic precipitators in Section 100.

In the gasifier which is operated around 105 kPa pressure and 1900°C temperature, lignite is partially oxidized to gases mainly containing CO, H_2, and CO_2. Because of high temperature, only traces of CH_4 (less than 0.1 vol.%) is formed. No other hydrocarbon gases such as ethane are formed. The sulfur is converted to H_2S (90%) and COS (10%). The raw gas after leaving the gasifier around 1500°C passes through a waste heat boiler unit which produces 3600 kPa saturated steam. Additional low-pressure (250 kPa) saturated steam is also generated by recovering the process heat in the reactor jacket of the gasifier.

Most of the lignite ash leaves the gasifier through the bottom in molten form and then is solidified by contact with water in a quench tank situated beneath the gasifier. The ash is removed from the quench tank by means of a scraper-conveyor assembly and is then conveyed to a slag storage bin which permits loading of the slag into railroad cars or trucks for disposal at the mine site or moved by belt conveyors to the mine site (depending upon the economic feasibility).

2.1.2 Coal Gasification (Section 200) (cont'd)

The raw gas from the waste heat boiler unit is cooled by direct water injection to solidify fly ash. Further removal of entrained particles is carried out in the electrostatic precipitators where the dust content of the raw gas is reduced to 0.2 mg/m^3.

The discharge water from the wash coolers is recirculated through a wash water system equipped with a clarifier to remove suspended solids. The clarifier overflow is pre-cooled through cooling tower before recirculation to the wash water system.

2.1.3 Gas treatment (Section 300)

In this section, the solids-free raw gas is compressed to about 3 MPa using steam-turbine-driven centrifugal compressors and is then scrubbed with methanol to remove HCN. The scrubbed gas is then treated in a Rectisol unit where sulfur-containing gases (H_2S and COS) are removed using cold sulfur-free methanol from a CO_2 stripper located downstream. The H_2S and COS feed, after methanol regeneration step, is sent to the Claus sulfur plant in Section 600.

About 56% of the desulfurized raw gas undergoes two-stage shift-conversion for preparation of synthesis gas. The shifted gas containing mainly CO_2 and H_2 is combined with remaining (unshifted) gas and fed to a Rectisol CO_2 removal unit where 90 vol.% of CO_2 is removed using methanol solution. The refrigeration demand for the Rectisol units is met by an ammonia-absorption refrigeration unit. The treated synthesis gas having the following mole ratio as required for methanol synthesis is then sent to Section 400.

$$\frac{H_2}{2CO + 3CO_2} = 1$$

2.1.4 Methanol Synthesis (Section 400)

For methanol synthesis, the ICI process is used. Synthesis gas from Section 300 is compressed to 5.3 MPa by multi-stage turbocompressor. This make-up synthesis gas then combines with the recycled unconverted synthesis gas from the gas-liquid separator of the reactor effluent. About 80% of the combined gas stream is preheated and then passed directly to the converter, which consists of several copper-zinc-aluminum oxide catalyst beds placed in series. The remaining 20% of the feed is introduced cold between the beds, in order to effect temperature control of the gas stream as it undergoes the following exothermic reactions:

$$CO + 2\,H_2 = CH_3OH$$

$$CO_2 + 3H_2 = CH_3OH + H_2O$$

The gas stream leaving the converter at about 270°C contains approximately 5% methanol, together with some light organic byproducts such as acetone and higher alcohols which are the result of undesirable side reactions. This exit stream first preheats the boiler feed water and the feed entering the converter is then water-cooled to partially condense methanol and some of the organic impurities. The condensate is separated from the gas stream. After pressure reduction (5 to 0.4 MPa), the dissolved gases are liberated from condensate in the flash drum. The flash gas is sent to a boiler in Section 900. Part of the flash gas is sent to the Shell-Claus-Offgas-Treating (SCOT) unit in Section 600. The crude methanol containing about 7 mole % water and 0.2 mole % dissolved gases (mainly CO_2 with traces of methane) is routed to storage which provides feed for methanol purification in Section 1100.

2.1.4 Methanol Synthesis (Section 400) (cont'd)

To avoid the inert gas build-up such as nitrogen in the feed gas mixture, a portion of the recycle synthesis gas is continuously purged and used as fuel gas in Section 900.

2.1.5 Oxygen Plant (Section 500)

High purity oxygen (98 mole %) required for lignite gasification is produced in a conventional medium pressure air-separation plant. Oxygen obtained from the plant is compressed and fed into the Koppers-Totzek gasifiers in Section 200.

High-purity nitrogen, which is also produced, is used for: transporting lignite in Section 100; inert blanketing of liquid product storage; stripping CO_2 from methanol in the CO_2 stripper (Section 300) and instrumentation purposes. Low-purity nitrogen is vented to atmosphere.

2.1.6 Emission Control (Section 600)

The acid-gas streams from the Rectisol unit (Section 300) are treated in the Claus sulfur plant for recovery of elemental sulfur.

About one-third of the total acid gas stream entering the Claus plant is oxidized with air in a furnace to convert the H_2S to SO_2. This SO_2-containing gas stream is combined with the remaining two-third acid-gas stream in a two-stage Claus converter to form elemental sulfur ($SO_2 + 2H_2S = 2H_2O + 3S$). The recovered molten sulfur is sent to a storage pit (Section 800) and the off-gases are further treated in a SCOT unit to reduce the sulfur content to an environmentally acceptable level.

2.1.6 Emission Control (Section 600) (cont'd)

In the SCOT unit, all sulfur compounds present in the tail-gas stream and fuel gases from Sections 100 and 900 are selectively converted to H_2S over a cobalt-molybdenum catalyst at 300°C in the presence of a reducing gas. The reducing gas used is the portion of expansion gas from Section 400 which contains sufficient quantity of CO and H_2. The exit gases from the SCOT unit, containing 2% to 3% H_2S, are treated in an amine absorber for removal of the H_2S impurity. Treated gases from the absorber are sent to the incenerator stack. Thus the SO_2 content in the vent gases is reduced to less than 130 ppm before being discharged to the atmosphere.

The spent amine is stripped to produce an H_2S-rich gas stream which is recycled to the Claus unit for sulfur recovery. The regenerated amine is recycled to the absorber.

2.1.7 Effluent Control (Section 700)

The waste water treatment facilities process wash water from gasification Section and waste water and blowdown streams from the entire plant to sufficient quality to be used as cooling tower (make-up) water. Sludge collected from the treatment system is thickened, aerobically digested and then concentrated on a gravity belt-filter before disposal.

2.1.8 Product Blending, Storage and Shipping (Section 800)

No product blending facilities are required for the methanol plant. Facilities for receiving and storage of raw materials and chemicals such as coal, caustic, etc., have been provided for a 30-day-production-run based on full-plant capacity.

2.1.8 Product Blending, Storage and Shipping (Section 800) (cont'd)

Coal will be stored as open stock-piles near the coal preparation area. Water supplied to the plant will be stored in a reservoir for 1-day maximum plant usage. This reservoir protects against a brief interruption of the raw water supply and also provides reserves for fire protection services.

All chemicals except caustic have been assumed to be delivered to the plant site in drums or bags and are stored in a warehouse. Caustic will be received by tankers (concentration 50°Be caustic), diluted to 20°Be during unloading, and stored in offsite tankage.

For intermediate products such as crude methanol, storage capacity for 2-day production has been assumed. For byproduct sulfur, storage capacity equivalent to 30-day production has been considered. Sulfur will be stored in a heated-pit and shipped in molten form at 130° to 145°C.

For pure methanol product, storage capacity of 5 days has been considered.

2.1.9 Utilities (Section 900)

The utilities service includes a full range of systems necesssary for the operation of the plant, such as raw water treatment, cooling water, boiler feed water and condensate water treatment, steam and electricity generation, plant and instrument air system.

River water is assumed to be the source of make-up water for the plant. This water will be treated in a cold lime unit and filtered to reduce hardness, carbonate alkalinity, turbidity and

2.1.9 **Utilities (Section 900)** (cont'd)

suspended organics. The softened water is used directly for cooling tower make-up and as process water. It is also supplied to the demineralization unit where cation, anion and mixed-bed ion exchangers are used for boiler feed water (BFW) preparation. The process condensate water from the entire plant is also treated in this unit. The demineralized water is sent to deaerators to remove any dissolved oxygen and CO_2. Low-pressure steam produced in the gasifier jacket is used. After addition of inhibitors, the treated boiler feed water is then distributed to onsite and offsite steam generators.

About 4% of the steam requirement for the entire plant is met by the heat recovery from the gasifier. The remaining steam is produced by coal-fired steam boilers. Superheating of the high-pressure steam is provided by firing fuel gas (expansion and purge gas from Section 400) in steam superheaters. The high-pressure superheated steam is used to drive backpressure turbines and recycle gas compressors. It is also used in extraction-type condensation turbines for electrical power generation.

Nitrogen from the air separation unit is distributed for inert gas blanketing, lignite conveying in Section 100, and for methanol stripping in Section 300.

2.1.10 Offsite Facilities (Section 1000)*

Offsite requirements include items such as administration buildings, laboratory, maintenance equipment and buildings, warehouses, a fire-house, roads and onsite railway trackage.

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* Not shown in overall black diagram (Drawing 2-1)

2.1.11 Product Upgrading (Crude Methanol Purification) (Section 1100)

The crude methanol contains about 75% methanol together with dimethyl ether, esters and ketones as well as dissolved synthesis recycle gases (CO, CO_2, H_2, CH_4 and N_2). It must be purified to 99.95 wt% in order to meet gasoline blending stock specifications.

The purification is carried out by extractive distillation using two fractionating columns. These columns operate at slightly over 100 kPa pressure, and both use steam-heated reboilers and water-cooled condensers. The impure methanol and a recycled water stream pass to the first column, where extractive distillation occurs, in which the water extracts the methanol and the volatile impurities are removed as overhead product. The uncondensed vapors mainly containing dissolved gases from the overheads condenser are sent to the boiler in Section 900. The aqueous mixture leaving the bottom of the first column is then separated from heavy-boiling impurities and water in the second column. The overhead stream containing methanol of 99.95 wt% purity is sent to methanol storage tanks in Section 800. A side stream containing the remaining higher alcohols is sent to the waste water treatment plant in Section 900. The bottom product stream mostly containing water is first used to preheat the incoming crude methanol and then is recycled to the first distillation column as the extracting agent for distillation.

The yield of pure methanol amounts to approximately 98.5% of the crude methanol fed to the distillation unit.

2.1.11 **Product Upgrading (Crude Methanol Purification) (Section 1100)** (cont'd)

In order to produce turbine fuel quality methanol (less than 2.0 wt% water), the process configuration will be identical to that described earlier for production of pure methanol (99.95 wt%). The only difference is that fewer trays are needed in both distillation columns to make 98 wt% methanol.

Table 2-2 Product Yield and Efficiencies in the Conceptual, Commercial Lignite-to-Methanol Plant

[Basis: 25,000 t/d "as-received" Coal]

Feed	Willow Bunch Lignite
Coal (As Received)	
Moisture, wt%	35.0
Ash, wt% of the MF coal	21.0
Coal Fed to Gasifier	
Moisture, wt%	8.0
Ash, wt% of MF coal	21.0
Input	
MF Coal to Gasifier, t/d	11,476
MF Coal to Offsites, t/d	4,774
Electricity, MW	-
Output, t/d (bbl/d)	
Methanol (99.95 wt%)	5,714 (43,761)
Sulfur	120
Electricity *, MW	2.0
Liquid Product Yield (based on coal to gasifier)	
bbl/t as-received coal	2.48
bbl/t MF coal	3.81
bbl/t MAF coal	4.83
Liquid Product Yield (based on coal to plant)	
bbl/t as-received coal	1.75
bbl/t MF coal	2.69
bbl/t MAF coal	3.41
Thermal Efficiency, %	
Overall	39.5
Liquid Products	39.1

* Excess Power

2.2 OVERALL MATERIAL BALANCE, THERMAL EFFICIENCY AND UTILITIES SUMMARY

The interrelationship between various processing areas and the overall material balance of a conceptual, integrated, self-sufficient, commercial coal-to-methanol plant is presented in Drawing 2-1. Product yield and thermal efficiencies are summarized in Table 2-2. The basis used is 25,000 t/d "as-received" Willow Bunch lignite to the plant, out of which 17,656 t/d is fed to the Koppers-Totzek gasifier. The remaining 7,344 t/d is used in the offsite boilers to meet the motive and process steam demands of the plant. The plant produces about 43,761 bbl/d of methanol having 99.95 wt% purity. This amounts to a yield of 2.48 bbl/t of "as-received" lignite to gasifier. On moisture-free (MF) and moisture-and-ash-free (MAF) basis, this corresponds to 3.81 bbl/t and 4.83 bbl/t respectively. The plant also produces 120 t/d of byproduct sulfur.

The utility balance of the entire plant, i.e., production and consumption of steam, boiler feed water, etc., is summarized in Table 2-3. The plant is completely self-sufficient in terms of steam and power requirements. About 49,800 t/d (7615 gpm) of make-up water is required from outside source. Since the plant design is based on the assumption that process streams are water cooled, the recycle cooling water volume is high (2,308,000 t/d). Consequently, the evaporation loss in the cooling tower is 33,600 t/d which represents 67% of the make-up water requirement. If the air cooling is employed wherever possible, the recycling cooling water volume, as estimated by Krupp-Koppers, will be reduced to about 960,000 t/d. Thus, the make-up water requirements will reduce from 49,800 t/d to approximately 16,800 t/d (2,569 gpm).

2.2 OVERALL MATERIAL BALANCE, THERMAL EFFICIENCY AND UTILITIES SUMMARY (cont'd)

The utility balances were developed on the basis of in-plant steam and power generation. However, a small surplus of electrical power (2 MW) is available for export which amounts to only 2% of the total power required to operate the plant.

The overall thermal efficiency of the plant is 39.5%. The efficiency is a measure of the overall conversion of coal energy to product energy. The input energy also includes energy of coal which is sent to offsite boilers to generate steam for the plant. The output energy includes product and byproduct energy as well as credit for surplus electricity.

The liquid product efficiency of the coal-to-methanol plant based on Koppers-Totzek gasifier is 39.1%. This indicates that 99% of the total output energy is contained in the methanol product.

Table 2-3 Summary of Utility Balance for an Integrated, Self-Sufficient, Commercial Coal-to-Methanol Plant [Basis: 25,000 t/d "As-Received" Willow Bunch Lignite]

UTILITY SECTIONS	Steam 10.8 MPa 500°C t/d	Steam 3.6 MPa 370°C t/d	Steam 3.6 MPa t/d	Steam 0.25 MPa t/d	Condensate Water * t/d	Boiler Feed Water t/d	Cooling Water t/d	Raw Water t/d	Electricity kWh/h
Lignite Preparation				288			4,800		18,800
Lignite Gasification						18,024	28,800		18,000
Gas Treatment		10,608	6,480	168			470,400		10,000
Methanol Synthesis	18,000						77,000		1,000
Oxygen Plant		11,088					228,000		700
Emission Control									950
Effluent Control	2,760			984					4,600
Product Blending									
Utilities	9,000					30,600	1,188,000		36,150
Offsites									1,950
Product Upgrading			5,640	3,960			211,000		850
TOTAL (CONSUMPTIONS)	29,760	21,696	12,120	5,400	–	48,624	2,208,000	–	93,000
Lignite Preparation					288				
Lignite Gasification			12,120	5,400					
Gas Treatment					13,416				
Methanol Synthesis		18,000							
Oxygen Plant					11,088				
Emission Control									
Effluent Control									
Product Blending									
Utilities	29,760	3,696			7,872		2,165,280	49,800	95,000
Offsites									
Product Upgrading									
TOTAL (PRODUCTION)	29,760	21,696	12,120	5,400	41,544	–	2,165,280	49,800**	95,000

* Includes Process Water

** Make-up water (includes 33,600 t/d for evaporation losses in cooling tower)

2.3 PRODUCT QUALITY

The composition of methanol product is 99.95 wt% methanol, 0.05 wt% water and traces of other alcohols. Methanol containing upto a 0.1 wt% maximum water meets the requirements for direct blending in gasoline.* Fuel grade methanol which may be used as turbine fuel etc., could contain upto 2 wt% water.

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* Neat methanol is referred to as methanol containing no water.

3.0 TECHNICAL ASSESSMENT OF A COAL-BASED, METHANOL-TO-GASOLINE PLANT (MOBIL-MTG)

3.0 TECHNICAL ASSESSMENT OF A COAL-BASED, METHANOL-TO-GASOLINE PLANT (MOBIL-MTG)

3.1 OVERALL CONFIGURATION OF A CONCEPTUAL, SELF-SUFFICIENT PLANT

The overall configuration of the coal-based, commercial methanol-to-gasoline commercial plant includes coal gasification by Koppers-Totzek entrained-flow process, ICI methanol synthesis unit and conversion of crude methanol to gasoline by Mobil-MTG process. Supporting systems such as oxygen plant, utilities supply and offsite facilities are also included in order to comprise an integrated, self-sufficient plant operation.

A list of processing areas or sections of a conceptual commercial plant is presented in Table 3-1. The interrelationship between these processing areas is shown in the overall block flow diagram, Drawing 3-1.

Most processing sections, particularly coal-to-crude-methanol synthesis (Section 100 to Section 700), are essentially identical to the coal-to-methanol plant configuration described earlier. Therefore, the following description concentrates on the methanol-to-gasoline conversion section (Product Upgrading - Section 1100). The process design of this section is based on data given by Mobil Research and Development Corporation in a U.S. Department of Energy report, entitled "Research Guidance Studies to Assess Gasoline From Coal by Methanol-to-Gasoline and SASOL-Type Fischer-Tropsch Technologies". The Mobil methanol-to-gasoline (MTG) conversion section includes crude methanol conversion unit followed by a fractionation and an alkylation units to yield synthetic gasoline.

3.1.1 Methanol Conversion (Section 1100)

The methanol conversion unit employs a fixed-bed reactor system packed with a very specific zeolitic type (crystalline, aluminosilicate) catalyst, developed by Mobil R&D corporation. The catalyst essentially dehydrates the methanol to the stoichiometric yield of hydrocarbons (44%) and water (56%). The generated hydrocarbons are predominantly in the gasoline boiling range (C_4 to C_{10}). The chemical composition of the gasoline is similar to conventional gasoline consisting of high-branched paraffins (50%), high-branched olefins (13%), naphthenes (8%) and aromatics (28%). Small amounts of CO, CO_2, coke and trace amounts of formic acid and acetone in the water product are formed as byproducts. Since the catalyst is tolerant to water and also converts other oxygenates, crude (or "wild") methanol without purification is supplied directly from the methanol synthesis plant (Section 400) to the MTG unit. The methanol conversion reaction is highly exothermic with heat of reaction being about 1700 kJ/kg of methanol. In the first step, methanol is partially dehydrogenated to form an equilibrium mixture of methanol, dimethyl ether and water. The reactor conditions at the inlet are about 2.1 MPa and 300°C. About 20% of the heat of reaction is released in this step without any recycle for reactor temperature control. In the second step, where hydrocarbon conversion takes place, a large recycle gas stream is used to limit the temperature rise across the catalyst bed to about 50°C.

After fourteen days on stream, enough coke accumulates on the catalyst to require regeneration. Therefore, out of six M-gasoline reactors installed, five are operated while the sixth undergoes regeneration. The regeneration is accomplished by circulating a controlled amount of oxygen in nitrogen and burning off the deposited coke at about 450°C.

Table 3-1 Processing Areas of a Conceptual, Commercial, Coal-Based Methanol-to-Gasoline Plant

SECTION NO.	DESCRIPTION
100	COAL PREPARATION
	Storage and Reclaiming
	Crushing and Drying
200	COAL GASIFICATION
	Lignite Distribution and Gasification
	Waste Heat Recovery
	Raw Gas Washing
	Slag Removal
300	GAS TREATMENT
	Raw Gas Compression
	Rectisol Sulfur Removal
	Shift Conversion
	Rectisol CO_2 Removal
400	METHANOL SYNTHESIS
500	OXYGEN PLANT
600	EMISSION CONTROL
	Claus Sulfur Recovery Plant
	SCOT Tail-Gas Processing Unit
700	EFFLUENT CONTROL
800	PRODUCT BLENDING, STORAGE AND SHIPPING*
900	UTILITIES
	Raw Water Treatment and Distribution
	Cooling Tower and Cooling Water System
	Boiler Feedwater Preparation
	Condensate Treatment
	Steam Generation and Distribution
	Electricity Generation and Distribution
	Fuel Gas Purification and Distribution
	Plant Air, Instrument Air and Inert Gas System
	Flare System
1000	OFFSITES *
1100	METHANOL-TO-GASOLINE (PRODUCT UPGRADING)
	Methanol Conversion
	Fractionation
	HF Alkylation

* Not shown in the block diagram (Drawing 3-1)

3.1.2 Fractionation

The purpose of this unit is to fractionate crude gasoline from MTG conversion unit in order to obtain a deethanized feed for alkylation (overheads) and a stabilized gasoline (bottoms).

3.1.3 Alkylation

The purpose of this unit is to increase the gasoline yield by the reaction between iso-butane and butenes/propenes. The reaction products are higher molecular weight iso-paraffins such as iso-octane which are blended with gasoline in the product blending (Section 800).

The design of this unit is based on the proprietary Phillips Petroleum HF alkylation process. The process utilizes hydrogen fluoride (HF) as a catalyst for the alkylation reaction. Commercial grade propane and butane (liquefied petroleum gases) are produced as byproducts from the alkylation unit.

3.1.4 Product Blending, Storage and Shipping (Section 800)

Product blending facilities are designed to provide continuous and automatic in-line blending of various components to yield a 10-RVP (Reid Vapor Pressure) gasoline. The design is similar to a typical commercial in-line blending operation.

Details for receiving and storage of raw materials and chemicals such as coal, caustic, di-isopropyl ether are similar to those described earlier in Subsection 2.1.8.

3.2 MATERIAL BALANCE

The material balance for Mobil-MTG plant operating with crude methanol produced from synthesis gas using Koppers-Totzek gasifier has been scaled from the MTG reactor yields reported by Mobil Research and Development Company (DOE Report FE-2447-13). These yields are presented in Table 3-2 and represent the best estimate for a commercial fixed-bed MTG reactor. As seen from Table 3-2, a methanol-to-water molal feed ratio of 13.28 is required. The crude methanol produced from synthesis gas using Koppers-Totzek gasififer contains 13.5 methanol-to-water mole ratio. Therefore, first, the methanol-to-water feed ratio is adjusted to 13.28 by adding water (0.16 moles per 100 moles of methanol). By adjusting the MTG reactor feed composition, the yield of each product per mole of methanol will then be the same as that yielded in the Mobil-MTG plant. From this point onwards, the material balance computations for the MTG conversion unit is the straight forward application of a scaling factor. However, care was taken to assure that only products which are produced in the MTG reactor are scaled.

The process flow diagrams for MTG conversion, fractionation and HF alkylation units are essentially the same as those presented in the Mobil study. These diagrams are attached in Appendix B along with the computed material balances for Willow Bunch lignite using the above-described methodology.

Table B-1 in Appendix B shows the material balance for the MTG conversion unit (referred to as Unit 150 in process flow diagram Drawing ZO-GEM-6915). The stream numbers shown in the Table B-1 correspond to the stream numbers shown in the process flow diagrams given in Appendix B. Streams without stream numbers are not included in this table. These are process utility streams and have been accounted for under plant utility balances.

Table 3-2 Mobil-MTG Fixed-Bed Reactor Yields

Components	Feed (moles)	Product (moles)
Coke (as $CH_{0.8}$)	0.000	0.128
Acetone	.000	.129
Formic Acid	.000	.163
Methanol	100.000	.000
Dimethylether	.000	.000
Water	7.529	106.951
Carbon Monoxide	.000	.017
Carbon Dioxide	.000	.053
Hydrogen	.000	.040
Methane	.000	.746
Ethane	.000	.193
Ethene	.000	.018
Propane	.000	1.474
Propene	.000	.064
n-Butane	.000	.665
i-Butane	.000	2.155
Butenes	.000	.270
n-Pentane	.000	.268
i-Pentanes	.000	2.353
Pentenes	.000	.435
Cyclopentane	.000	.047
Methylcyclopentane	.000	.214
n-Hexane	.000	.113
i-Hexanes	.000	1.991
Hexenes	.000	.297
Methylcyclohexane	.000	.062
n-Heptane	.000	.026
i-Heptanes	.000	.786
Heptenes	.000	.292
1,3 Dicyclopentane, cis	.000	.233
1-Octenes	.000	.228
Octenes	.000	.301
n-Propylcyclopentane	.000	.299
n-Nonane	.000	.015
i-Nonanes	.000	.084
Nonenes	.000	.116
n-Butylcyclopentane	.000	.071
i-Decanes	.000	.024
Decenes	.000	.045
Benzene	.000	.036
Toluene	.000	.280
Ethylbenzene	.000	.070
m- + p-Xylenes	.000	.876
o-Xylene	.000	.240
1,2,4-Trimethylbenzene	.000	.818
1,3,5-Trimethylbenzene	.000	.034

Table 3-2 Mobil-MTG Fixed-bed Reactor Yields (cont'd)

Components	Feed (moles)	Product (moles)
p-Ethyltoluene	0.000	0.292
i-Propylbenzene	.000	.014
1,2,4,5-Tetramethylbenzene	.000	.436
1,2,3,5-Tetramethylbenzene	.000	.063
1,2,3,4-Tetramethylbenzene	.000	.023
p-Diethylbenzene	.000	.198
Penta-Methylbenzene	.000	.068
2-Methylnaphthalene	.000	.017
Total moles	107.529	124.832

Source: DOE Report FE-2447-13, Aug. 1978

3.2 MATERIAL BALANCE (cont'd)

Table B-2 in Appendix B shows the material balance for Fractionation Unit (Unit 152 in the process flow diagram - Drawing ZO-GEM-6916). It should be noted that since there is no SNG train in the plant configuration shown in Drawing 3-1 (because no methane is produced in the Koppers-Totzek gasifier), the material flow rate of stream 58 is equal to zero. Stream 57 (light gas to SNG train) in the process flow diagram is not methanated because of insufficient hydrogen. Therefore, it has been considered as fuel gas product. Alternatively, this fuel gas could be utilized in the offsites to raise steam and thus, the amount of coal required in offsites will be reduced.

The material balance of HF alkylation unit (Unit 153 in process flow diagram - Drawing ZO-GEM-C916) is presented in Table B-3 in Appendix B.

3.3 OVERALL MATERIAL BALANCE, PRODUCT EFFICIENCIES AND UTILITY SUMMARY

The interrelationship between various processing areas and the overall material balance of an conceptual, integrated, self-sufficient, commercial, coal-based methanol-to-gasoline plant is presented in Drawing 3-1. Product yield and thermal efficiencies are summarized in Table 3-3. The basis used is 25,000 t/d of "as-received" Willow Bunch lignite to the plant out of which 17,656 t/d is fed to the Koppers-Totzek gasifier. The remaining 7,344 t/d is used in the offsite boilers to produce steam and electricity. The overall gasoline product yield is about 1.1 bbl/t of "as-received" lignite to gasifier. This corresponds to about 1.7 bbl/t and 2.1 bbl/t of lignite on moisture-free (MF) and moisture-and-ash-free (MAF) basis respectively. The byproducts are 64 t/d fuel gas (GCV = 37,643 kJ/kg), 117 t/d propane and 186 t/d butane as LPG. About 120 t/d of sulfur is also produced as byproduct.

Product yield and thermal efficiency may be improved by process optimization which is out of the scope of this study. Process heat recovery improvements, use of gas products as fuel for offsites, etc., are some of the possible ways that may be considered. As an example, in Table 3-3, the product yields and thermal efficiency are presented when fuel gas product is used in offsites. In this case, the amount of coal (MF) required in offsites decreases from 4,774 to 3,410 t/d. Thus, coal input to the gasifier increases by 1,364 t/d (MF basis). This results in about 12% increase in the product yield (propane, butane and gasoline).

The utility balance of the entire plant, i.e., production and consumption of steam, boiler feed water, etc., is summarized in Table 3-4. The plant is self-sufficient in terms of steam and

Table 3-3 Product Yield and Efficiencies of a Conceptual, Commercial Coal-Based, Methanol-to-Gasoline Plant

[Basis: 25,000 t/d "as-received" coal]

Feed	Willow Bunch Lignite (Fuel Gas as Product)	Willow Bunch Lignite (Fuel Gas to Offsite)
Coal (as received)		
Moisture wt%	35.0	35.0
Ash, wt% of the MF coal	21.0	21.0
Coal Fed to Gasifier		
Moisture, wt%	8.0	8.0
Ash, wt% of MF coal	21.0	21.0
Input		
MF Coal to Gasifier, t/d	11,476	12,840
MF Coal to Offsites, t/d	4,774	3,410
Fuel Gas to Offsites, t/d	-	64
Electricity, MW	-	-
Output, t/d (bbl/d)		
Fuel Gas	64	-
Propane (LPG)	117 (1456)	131 (1624)
Butane (LPG)	186 (2062)	208 (2307)
Gasoline	2,245 (19285)	2,512 (21577)
Sulfur	120	120
Liquid Product Yield (based on coal to gasifier)		
bbl/t as-received coal	1.09	1.09
bbl/t MF coal	1.68	1.68
bbl/t MAF coal	2.13	2.13
Liquid Product Yield (based on coal to plant)		
bbl/t as-received coal	0.77	0.86
bbl/t MF coal	1.19	1.33
bbl/t MAF coal	1.50	1.68
Thermal Efficiency, %		
Overall	43.5	40.5
Liquid Products	31.4	35.1

Table 3-4 Summary of Utility Balance of a Self-Sufficient, Commercial Coal-Based Methanol-to-Gasoline Plant

[Basis: 25,000 t/d "As-Received" Willow Bunch Lignite]

UTILITY / SECTIONS	Steam 10.8 MPa 500°C t/d	Steam 3.6 MPa 370°C t/d	Steam 3.6 MPa t/d	Steam 0.25 MPa t/d	Condensate Water * t/d	Boiler Feed Water t/d	Cooling Water t/d	Raw Water t/d	Electricity kWh/h
Lignite Preparation				288			4,800		18,800
Lignite Gasification						18,024	28,800		18,000
Gas Treatment		10,608	6,480	168			470,400		10,000
Methanol Synthesis	18,000						96,000		1,000
Oxygen Plant		11,088					228,000		700
Emission Control				144			8,880		1,180
Effluent Control	2,760			984					8,800
Product Blending									200
Utilities	9,000		4,800			30,600	1,188,000		36,150
Offsites									1,950
Product Upgrading			840	72		648	4,320		2,020
TOTAL (CONSUMPTIONS)	29,760	21,696	12,120	1,656		49,272	2,029,200		98,800
Lignite Preparation					288				
Lignite Gasification			12,120	5,400					
Gas Treatment					13,416				
Methanol Synthesis		18,000							
Oxygen Plant					11,088				
Emission Control									
Effluent Control									
Product Blending									
Utilities	29,760	3,696			17,064		1,986,480	50,160**	99,000
Offsites									
Product Upgrading				648	4,368				
TOTAL (PRODUCTION)	29,760	21,696	12,120	6,048	46,224		1,986,480	50,160	99,000

* Includes Process Water
** Make-up water (includes 42,700 t/d for evaporation losses in cooling tower)

3.3 OVERALL MATERIAL BALANCE, PRODUCT EFFICIENCIES AND UTILITY SUMMARY (cont'd)

power requirements. The electrical power consumption is about 99 MW which is generated in the offsites. This power consumption is based on the plant configuration where steam turbines are used for power intensive operations such as raw gas and recycle gas compression. About 50,160 t/d (7,364 gpm) of make-up water is required from outside source. As discussed earlier in Section 2-2, if the air cooling is used for process streams wherever possible, the make-up water requirements will decrease to approximately 16,800 t/d (2,467 gpm).

The overall thermal efficiency of the plant is 43.5%. The input energy includes coal to gasifiers for synthesis gas production and to offsite boilers to generate steam and power for the plant. The output energy includes product and byproduct energy.

The liquid product (i.e., gasoline) efficiency of the coal-based methanol-to-gasoline plant operating with Willow Bunch lignite is 31.4%. This means about 72.2% of total output energy is contained in the gasoline product. If propane and butane (LPG) are also included, the liquid products efficiency increases to 35.9%.

In the case where fuel gas is utilized in offsites, the overall thermal efficiency decreases from 43.5% to 40.5%, but the liquid product efficiency increases from 31.4% to 35.1%.

3.4 PRODUCT QUALITY

3.4.1 Gasoline

The composition of blended gasoline product from the Mobil-MTG (Methanol-to-Gasoline) process is given below:

Composition	wt%
Mixed Butanes	2.6
Alkylate	3.6
Stabilized Gasoline	94.0
	100.0

Estimated properties and a comparison with unleaded gasoline specifications are presented in Table 3-5. The 10-RVP gasoline is a premium quality motor fuel and meets all specifications. Its octane rating is estimated to be about one above the specification number. The volatility and PONA (paraffins, olefins, naphthenes and aromatics) composition are similar to present-day gasoline. It should be noted that although the durene content is below the target maximum of 5.0 wt%, it would be preferable if it were below 4.0 wt%.

3.4.2 Propane and Butane LPG

Estimated properties of Mobil-MTG propane and butane and a comparison with LPG specifications are shown in Table 3-6. These products meet commercial propane LPG and butane LPG specifications.

Table 3-5 Comparison of Principal Unleaded Gasoline Specification with Estimated MTG Gasoline Properties

	Estimated MTG Unleaded Gasoline Properties	Specification
Gravity, °API	61.4	
Octane Numbers		
Research	93	
Motor	83	82 min.
(Research + Motor)/2	88	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation, °C		
IBP	29.7	
10%	43.6	70 max.
30%	63.4	
50%	93.3	767/121 min./max.
70%	128.2	
90%	169.4	190 max.
EP	198.1	225 max.
V/L Ratio (=20), °C	@ 54.2	@ 60 max.
Sulfur, wt%	Nil	0.10 max.
Composition, vol%		
Paraffins	51.3	
Olefins	11.3	20 max. (target)
Naphthenes	8.7	
Aromatics	28.7	
Durene Content, wt%	4.6	5 max. (target)
Molecular Weight	94.3	

Source: DOE Report FE-2447-9, Jan. 1978

Table 3-6 Comparison of Principal LPG Specifications with Estimated Methanol-To-Gasoline LPG Properties

	Commercial Propane ASTM D1835*	Est. MTG C_3 LPG Properties	Commercial Butane ASTM D1835*	Est. MTG C_4 LPG Properties
Vapor Pressure at 37.8°C, max. psig (D1267)	208	171	70	52
Volatile Residue				
Butane and Heavier, max. vol %	2.5	1.9	–	–
Pentane and Heavier, max. vol %	–	–	2.0	0.7
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

* 1976 Annual Book of ASTM Standards

Source: DOE Report FE-2447-9, Jan. 1978

4.0 TECHNICAL ASSESSMENT OF A COAL-BASED, SASOL-TYPE, FISCHER-TROPSCH SYNTHESIS PLANT

4.0 TECHNICAL ASSESSMENT OF A COAL-BASED, SASOL-TYPE, FISCHER-TROPSCH SYNTHESIS PLANT

4.1 OVERALL CONFIGURATION OF A CONCEPTUAL, SELF-SUFFICIENT PLANT

The overall configuration of the commercial coal-to-gasoline plant includes coal gasification by Kopper-Totzek entrained-flow process and syngas synthesis using SASOL-type Fischer-Tropsch technology to produce gasoline. Supporting systems such as oxygen plant, utilities supply and offsite facilities are also included to comprise an integrated, self-sufficient plant operation.

A list of processing areas or sections of a conceptual commercial plant is presented in Table 4-1 and the interrelationship between these processing areas is shown in the overall block flow diagram in Drawing 4-1. An overall material balance for main process streams around each processing section is also included in the Drawing 4-1.

Many of the processing sections particularly coal preparation, coal gasification, gas treatment, oxygen plant, emission and effluent control, are essentially identical to the coal-to-methanol plant configuration described earlier in Section 2.1. Only material flow rates are changed. Therefore, the following description concentrates on Fischer-Tropsch Synthesis (Section 200A) and methanation (Section 400).

4.1.1 Fischer-Tropsch Synthesis (Section 200A)

The Fischer-Tropsch (F-T) synthesis of syngas is based on SASOL (South Africa Coal, Oil and Gas Corporation, Ltd.) technology

Table 4-1 Processing Areas of a Coal-Based, SASOL-type, Commercial Plant via Fischer-Tropsch Synthesis

SECTION NO.	DESCRIPTION
100	COAL PREPARATION Storage and Reclaiming Crushing and Drying
200	COAL GASIFICATION Lignite Distribution and Gasification Waste Heat Recovery Raw Gas Washing Slag Removal
200A	FISCHER-TROPSCH (F-T) SYNTHESIS Synthol F-T Synthesis F-T Fractionation F-T Product Hydrotreating Hydrotreater Product Fractionation Catalytic Reforming C_5/C_6 Isomerization Catalytic Polymerization HF Alkylation Polymer Gasoline Hydrogenation Light Ends Recovery Hydrogen Purification Alcohol Recovery Hydrocarbon Recovery Hydrogen Recovery
300	GAS TREATMENT Raw Gas Compression Rectisol Sulfur Removal Shift Conversion Rectisol CO_2 Removal
400	METHANATION CO_2 Removal SNG Drying
500	OXYGEN PLANT
600	EMISSION CONTROL Claus Sulfur Recovery Plant SCOT Tail-Gas Processing Unit

Table 4-1 Processing Areas of a Coal-Based, SASOL-type, Commercial Plant via Fischer-Tropsch Synthesis (cont'd)

SECTION NO.	DESCRIPTION
700	EFFLUENT CONTROL
800	PRODUCT BLENDING AND/OR STORAGE AND SHIPPING*
900	UTILITIES Raw Water Treatment and Distribution Cooling Tower and Cooling Water System Boiler Feedwater Preparation Condensate Treatment Steam Generation and Distribution Fuel Gas Purification and Distribution Plant-Air, Instrument Air and Inert Gas System Flare System
1000	OFFSITES *

* Not shown in the block diagram (Drawing 4-1)

4.1.1 Fisher-Tropsch Synthesis (Section 200A) (cont'd)

which employs dry-ash Lurgi gasifiers of German design and fast-fluid-bed (entrained recirculating) Synthol F-T reactors developed by SASOL (South Africa) and M.W. Kellogg, (U.S.A.). As explained in Chapter 1, major disadvantages with Lurgi gasifiers are the high quantities of methane they produce, and their inability to process coal fines. On the other hand, the Koppers-Totzek gasifier was judged suitable for Willow Bunch lignite. Therefore, the process configuration presented in the F-T Synthesis (Section 200A) is based on Koppers-Totzek gasifier and Synthol F-T reactor. Much of the detailed technical data on SASOL is proprietary. SNC has therefore employed a detailed technical description of a SASOL-type conceptual plant by Mobil Research and Development Corporation (MRDC) and Lurgi, based on information released by SASOL, and presented in DOE Report FE 2447-13.

The Fischer-Tropsch synthesis route requires thirteen major processing steps to achieve marketable synthetic fuels. These are listed under Section 200A in Table 4-1. A brief description of the processing technology involved in each step is presented below.

4.1.1.1 Synthol F-T Synthesis

Synthesis gas from Section 300 having a hydrogen-to-carbon monoxide (H_2/CO) molal feed ratio of 2.54 as required for the Synthol reactor is compressed to 2.2 MPa pressure and preheated to about 160°C temperature before it is fed to the reactor. Figure 4-1 shows a schematic of the operation of the Synthol reactor system. It can be described as a circulating fluidized catalyst system. The incoming synthesis gas carries the speci-

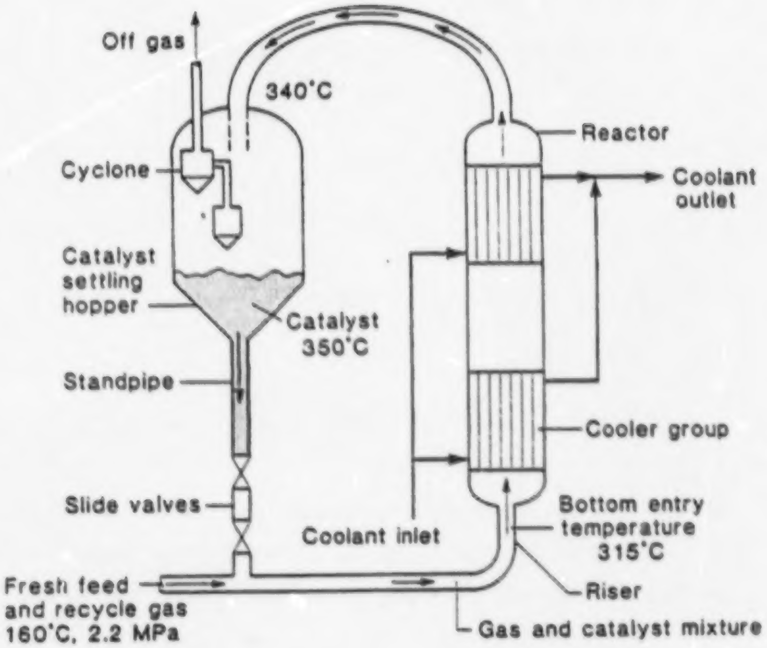

Figure 4-1 Schematic Diagram of Synthol Fischer-Tropsch Reactor

Source: J.C. Hoogendoorn, "Experience with Fischer-Tropsch Synthesis at SASOL", IGT conference, Chicago, Sept. 1973 fically prepared iron catalyst upwards through the reactor at a rapid rate. Approximately 7300 t/h of catalyst is circulated through the reactor. The catalyst preparation is described in Appendix C. The reactor products are a mixture of hydrocarbons (paraffins, mono-olefins and aromatics), alcohols, aldehydes, ketones and fatty acids. The following overall equations illustrate the F-T reactions:

$$nCO + 2n\,H_2 \longrightarrow \underset{\text{Hydrocarbon}}{(-CH_2-)\,n} + n\,H_2O$$

$$2nCO + nH_2 \longrightarrow \underset{\text{Hydrocarbon}}{(-\,CH_2\,-)\,n} + nCO_2$$

$$nCO + 2n\,H_2 \longrightarrow \underset{\text{Alcohol}}{H\,(-\,CH_2-)\,n\,OH} + (n-1)\,H_2O$$

$$CO + H_2O \longrightarrow H_2 + CO_2 \quad \text{(water gas shift)}$$

These equations do not indicate the reaction mechanism but they do show the stoichiometric relationship between reactants and products. Except for the water gas shift reaction, all reactions are highly exothermic. Most of the exothermic heat is removed by two sets of coolers.

The product gas mixture exits the reactor at about 350°C and is separated by cyclones from the finely ground catalyst in the disengaging section. The catalyst settles out into a hopper and flows down the standpipe to be swept back into the reactor by incoming synthesis gas. The catalyst life is about 45 days.

4.1.1.1 Synthol F-T Synthesis (cont'd)

Because of the expected fouling nature of the reactor effluent, it is water-washed and cooled. The water-washed stream containing alcohols and acids is sent to the alcohol recovery unit. A hot heavy-oil stream is withdrawn from the bottom of wash tower and sent to the F-T fractionation unit. The overhead vapors are cooled, condensed and water washed. The uncondensed vapor stream is split into recycle and purge gas streams. The latter is sent to the hydrocarbon recovery unit. The condensed light-oil stream is sent to the F-T product fractionator. Most of the process heat is recovered by generating low-pressure steam.

4.1.1.2 F-T Product Fractionation

The F-T product fractionator splits the main raw product stream (light- and heavy-oil streams) into four streams. The light ends containing mostly C_2 are sent to the hydrocarbon recovery unit. The bulk of the liquid is sent for hydrotreatment in the F-T product hydrotreating unit. A C_3/C_4 olefins stream is sent to the catalytic polymerization unit. The heavy fuel oils and residuals (450°C^+ residuum) are sent to the offsite boilers for firing.

4.1.1.3 F-T Product Hydrotreating

The purpose of this unit is to saturate the olefins and to eliminate the remaining alcohols and acids. The hydrogen from the hydrogen recovery unit is combined with the olefins feed and is then vaporized by heat exchange with the hydrotreater effluent stream. Additional heat required for vaporization is supplied by fired heaters. The feed then passes through the reactor where

4.1.1.3 F-T Product Hydrotreating (cont'd)

olefins are saturated, and sulfur and nitrogen are hydrogenated to form hydrogen sulfide and ammonia. The hydrotreated stream is then flashed through a two-stage flash system. Liquid products are combined and sent to a fractionation unit. The hydrogen-rich gases are sent to the polymer gasoline hydrogenation-unit. Reaction water is collected and sent to effluent control (Section 700).

4.1.1.4 Hydrotreater Product Fractionation

The purpose of this unit is to separate the hydrotreated product into a pentane/hexane stream (for isomerization), naphtha (for reforming), a diesel oil product and a heavy fuel oil product.

After heating, the hydrotreater product is fed to the main fractionator of a conventional three-tower system unit. The heavy fuel oil product is withdrawn from the bottom. A small amount of the heavy fuel oil stream (about 6%) has been used in the adjustment of the plant's fuel balance. A side-cut stream is withdrawn to steam stripper from which the diesel oil product is obtained. The light gases, butanes, pentanes/hexanes and naphtha in the overhead product are cooled and separated. Light gases become feed to the light-ends recovery unit. The heavy naphtha is separated from butanes in the debutanizer and then split into a pentane/hexane stream. Heavy naphtha from the bottom of the splitter is sent for reforming.

4.1.1.5 Catalytic Reforming

The purpose of catalytic reforming is to increase the anti-knock quality of the naphtha. It is carried out using platinum-based

4.1.1.5 Catalytic Reforming (cont'd)

platinum-based catalysts and the reactions include: (i) dehydrogenation of naphthenes; (ii) cyclization of straight-chain paraffins and subsequent dehydrogenation; (iii) isomerization of paraffins and cyclopentane derivatives and; (iv) hydrocracking of paraffins and naphthenes. Commercial processes licensed by Union Oil Products (UOP), Exxon, Chevron or Standard Oil may be used.

The catalytic reforming unit produces a reformate stream which is sent to gasoline blending in Section 800 and a C_3/C_4 stream sent to the light-ends recovery unit.

4.1.1.6 C_5/C_6 Isomerization

The light fraction containing C_5 and C_6 components is catalytically isomerized using platinum-containing catalyst. Isomerization increases the anti-knock quality of the gasoline. In order to maximize the octane improvement, the isomerization unit includes n-pentane recycling.

The C_5/C_6 product stream from the hydrotreater product fractionation unit along with the recycled n-pentane stream is first fed to a deisopentanizer column. The overhead from this column containing pentane isomerate, butanes and iso-pentane is cooled and directly sent to gasoline blending in Section 800.

The bottom stream is mixed with recycle and make-up hydrogen, pre-heated and then passed over the catalyst bed of the isomerization reactor. The reactor effluents are cooled and flashed. The condensed liquid is fractionated to separate light ends from hexane isomerate before it is sent to gasoline blending. The uncondensed hydrogen-rich gas stream after flashing is sent to the hydrogen purification unit.

4.1.1.7 Catalytic Polymerization

Catalytic polymerization is necessary for refining Fischer-Tropsch reaction products because of the high concentration of propene and butene olefins produced. The polymerization is catalyzed by acids and is generally an 85% conversion process yielding trimers with propene feeds and dimers with butenes. These polymerized products are blended into gasoline.

The water content of the feed is critical depending upon reactor operating conditions. The feed is first saturated with water before being fed to the inlet of polymerization reactor. The reactor effluent is fed to a debutanizer. The overhead stream from the debutanizer consisting of unreacted olefins, iso-butane and inert propane and n-butane is cooled and pumped to the HF alkylation unit. The bottom stream is split into a light gasoline stream and a C_9-C_{10} boiling range heavy polymer gasoline stream. The light gasoline is pumped to product blending (Section 800). The heavy gasoline is sent to the gasoline hydrogenation unit.

4.1.1.8 HF Alkylation

Iso-butane and unpolymerized C_3/C_4 olefins are fed to the HF catalyzed alkylation unit to increase the gasoline yield. The resulting alkylate containing higher molecular weight iso-paraffin is sent to product blending (Section 800). Propane and butane (LPG) gases are also produced in this unit.

4.1.1.9 Polymer Gasoline Hydrogenation

The olefin target concentration for the 10-RVP (Reid Vapor Pressure) gasoline product is 20 vol% maximum. Therefore, in

4.1.1.9 Polymer Gasoline Hydrogenation (cont'd)

order to saturate the olefins, heavy polymer gasoline is hydrogenated.

The heavy gasoline from catalytic polymerization unit is mixed with make-up and recycle hydrogen before it is fed to the reactor. The reactor uses multi-beds with hydrogen quench for temperature control. Reactor effluents are cooled and flashed. Uncondensed gases are sent to the hydrogen recovery unit. The condensed stream after stabilization is sent to gasoline blending (Section 800).

4.1.1.10 Light Ends Recovery

In the light ends recovery unit, light gasoline and iso-butane fractions are separated from the hydrocarbon-rich offgases produced in the hydrotreater product fractionation, catalytic reforming and C_5/C_6 isomerization units. The combined offgas stream from these units is first fed to a deethanizer. The deethanizer bottoms are then fed to a debutanizer which yields an overhead iso-butane stream and a light gasoline product stream as the bottoms. Iso-butane stream is sent to the HF alkylation unit. The uncondensed deethanizer overhead stream is utilized as a fuel for the plant.

4.1.1.11 Hydrogen Purification

This unit is based on commercial cryogenic technology. It provides make-up hydrogen stream of 90% purity from hydrogen-rich offgases produced in the catalytic reforming, C_5/C_6 isomerization, gasoline hydrogenation units.

4.1.1.12 Alcohol Recovery

The main purpose of this unit is to produce a marketable alcohol mixture product from the alcohol/acid water stream produced in the Fischer-Tropsch synthesis unit. Distillation is used to separate the bulk of the water and acids from raw alcohols. Methanol is then fractionated from the alcohol mixture. A benzene azeotropic system is used for final water removal. The final step incorporates hydrogenation of the aldehydes and ketones to improve the marketing value of the alcohol mixture. Make-up hydrogen is supplied from the hydrogen recovery unit.

4.1.1.13 Hydrocarbon Recovery

Heavy hydrocarbons carried over in the Fischer-Tropsch synthesis purge gas are recovered in this unit. The resulting product gas is then sent for methanation (Section 400).

The unit consists of a low temperature (-43°C) heptane wash system. This operating temperature is high enough to prevent the freezing of CO_2 present in the feed. The feed gas is first cooled against product gas and then by refrigeration to condense most of the higher hydrocarbons before being fed to a wash absorber. In the wash absorber, feed gas flows countercurrently to the heptane solvent. The rich heptane leaving the absorber is fed to the heptane regenerator in which the absorbed hydrocarbons are removed by fractionation. The recovered hydrocarbon stream is pumped to the F-T product fractionation unit. The absorber overhead product gas is sent for methanation (Section 400).

4.1.1.14 Hydrogen Recovery

This unit recovers hydrogen for use in the plant hydrogenation and hydrotreating units. The feedstock is a slip stream of the product gas from the hydrocarbon recovery unit. Because of relatively high CO_2 content (20 wt%) of the feed gas, the Pressure Swing Adsorption process of Union Carbide Corporation, New Jersey, is used. The unit consists of pressure-swing adsorbers where essentially all of the non-hydrogen components and some hydrogen are absorbed. The remaining hydrogen leaves the adsorber at a purity of over 98%. The adsorbed gases after desorption during regeneration cycle are sent to Section 400 to provide hydrogen for methanation reactions.

4.1.2 Methanation (Section 400)

The purpose of this unit is to reduce the CO content of feed gas from 2.6 vol% to less than 0.1 vol%. Using the Lurgi methanation process, feed gases from hydrogen recovery and fractionation units along with hydrogen-rich offgas from the hydrogen purification unit are combined and fed in series to the two downflow, fixed-bed methanation reactors. Methanated gas from first reactor flows to the second or clean-up reactor in which final methanation takes place. After recovering the process heat by generating MP steam, the gases are treated for CO_2 removal. The CO_2 removal is carried out by using the Amine Guard Process of Union Carbide. The purified gas is finally compressed and dried using the conventional triethyl glycol (TEG) wash process commercially used for drying of natural gas. Because methane is not produced in the gasifier, the final composition of methanated gas is 50.0 mole % H_2, 48.5 mole % CH_4 and traces of CO_2 and N_2.

4.1.3 Product Blending, Storage and Shipping (Section 800)

Details of product blending storage and shipping are similar to those described previously (Subsection 3.1.4). The storage facilities for intermediates include 5 days for hydrotreater feed and 2 days for reformer and catalytic polymerization unit feeds. Gasoline, diesel oil and heavy fuel oil product storage of 5 days have been considered.

4.2 MATERIAL BALANCE

The material balance for the Synthol Fischer-Tropsch synthesis plant (Section 200A in Drawing 4-1) using synthesis gas from Koppers-Totzek gasifier has been scaled from the Synthol reactor yields achieved in the SASOL base case reported in DOE Report FE-2447-13. Synthesis gas from the Koppers-Totzek gasifier is shifted to the H_2/CO ratio of 2.54 required in the Synthol reactor. It is [illegible] assumed that the yield of each product per mole of $H_2 + CO$ in the synthesis gas will be the same as that yielded in the SASOL base case. It should be noted that in the absence of actual experimental Synthol reactor yield data obtained from the synthesis gas from the Koppers-Totzek gasifier, the approach used to compute reactor yields provides the best estimates of the product yield.

The material balance computations are then based on the application of a scaling factor. However, care must be taken to ensure that only products which are produced from synthesis gas are scaled. SNG derived from methane, ethane and ethylene in the gasifier output, and gasoline derived from naphtha must be separately accounted for by subtracting them from the base-case output before the scaling factor is applied, since these components of Lurgi-derived synthesis gas are practically absent in the synthesis gas obtained from the Koppers-Totzek gasifier. Appropriate values for methane produced in the F-T synthesis are then accounted for in the methanation unit (Section 400).

Table 4-2 shows feed gas composition in the SASOL base case (Lurgi gasifier) and the Saskoil case (Koppers-Totzek gasifier). Table 4-3 presents the computed Synthol reactor yields for Saskoil case along with the base-case yields.

Table 4-2 Fischer-Tropsch Synthesis Feed Gas Composition

Feed Gas Composition, mole %	SASOL Base Case*	Saskoil Case (Willow Bunch Lignite)
CO_2	3.56	0.54
C_2H_4	0.03	-
CO	22.71	27.72
H_2	57.68	70.62
CH_4	15.30	0.06
C_2H_6	0.34	-
N_2 + Ar	0.38	1.06
	100.0	100.0

* Source: DOE Report FE-2447-13, Aug. 1978

Table 4-3 Fischer-Tropsch Synthol Reactor Yields

Component	Molecular Weight	Yields, wt% SASOL Base Case**	Yields, wt% Saskoil Case***
N_2	28.02	0.91	3.04
CO_2*	44.01	14.28	5.51
CO *	28.01	1.42	0.58
H_2	2.01	1.96	3.00
C_1	16.04	23.88	4.55
C_2^-	28.05	1.11	1.55
C_2	30.07	2.57	2.51
C_3	42.08	3.27	4.84
C_3	44.09	0.57	0.85
C_4^-	56.10	2.21	3.44
$i\text{-}C_4^-$	56.10	0.13	0.02
C_4	58.12	0.31	0.46
$i\text{-}C_5^-$	70.13	1.19	1.77
iC_5	70.13	0.50	0.74
C_5	72.15	0.30	0.44
$i\text{-}C_6^-$	84.16	0.84	1.24
$i\text{-}C_6$	84.16	0.50	0.74
C_6	86.17	0.33	0.49
200	97	1.03	1.52
250	110	1.60	2.36
300	127/122	0.93	1.36
350	144	1.39	2.06
400	165/163	0.77	1.14
450	187	1.03	1.51
500	211/210	0.48	0.71
550	238/232	0.63	0.94
600	260	0.23	0.34
650	288	0.18	0.26
700	315	0.31	0.45
750	350	0.23	0.34
800	380	0.15	0.22
850	409	0.12	0.17
900	440	0.04	0.06
1000^+	510	0.15	0.23
Alcohols	51.34	2.48	3.66
Acids	65.26	0.53	0.79
H_2O*	18.01	31.44	45.84
		100.00	100.00

* Yields reflect shift equation equilibrium
** Data from DOE Report FE-2447-13, Aug. 1978
*** Yields for Willow Bunch lignite computed from SASOL base case

4.3 OVERALL MATERIAL BALANCE, THERMAL EFFICIENCY AND UTILITY SUMMARY

The overall material balance of a conceptual, self-sufficient, commercial plant based on Fischer-Tropsch synthesis to produce gasoline as the major product is presented in the overall block diagram, Drawing 4-1. Product yields and thermal efficiencies are presented in Table 4-4. The basis used is 25,000 t/d of "as-received" Willow Bunch lignite to the plant. About 71.5% of this lignite is fed to the Koppers-Totzek gasifier to produce synthesis gas. The remaining is used in the offsite boiler to produce steam and electricity.

The overall product yield includes gasoline, diesel oil and fuel oil as main liquid fuel products. The other products are high calorific value fuel gas, propane and butane as LPG and mixed alcohols. Sulfur is a byproduct.

The overall liquid product yield is about 0.65 bbl/t of "as-received" lignite to gasifier. This corresponds to about 1.0 bbl/t and 1.26 bbl/t of lignite on moisture-free (MF) and moisture-and-ash-free (MAF) basis respectively.

Table 4-5 presents the comparison of overall product yields between the "base" case and the Saskoil case. On the basis of 1000 tonnes MAF coal to the gasifier, 8.5 tonnes of additional liquid products (gasoline, diesel and fuel oil) and 1.3 tonnes of additional propane and butane as LPG is produced in the Saskoil case when compared with the base case. On the other hand, in the Saskoil case, 163 tonnes less of SNG is produced. These differences are primarily due to the different composition of synthesis gas (see Table 4-2) produced by Lurgi and Koppers-Totzek gasifiers.

The utility balance of the entire plant, i.e., production and consumption of steam, boiler feed water, electrical power, etc.,

Table 4-4 Product Yield and Efficiencies of a SASOL-Type Plant via Fischer-Tropsch Synthesis (Synthol Reactor)

[Basis: 25,000 t/d "as-received" Lignite]

Feed	Willow Bunch Lignite
Coal (as received)	
Moisture, wt%	35.0
Ash, wt% of the MF coal	21.0
Coal Fed to Gasifier	
Moisture, wt%	8.0
Ash, wt% of MF coal	21.0
Input	
MF Coal to Gasifier, t/d	11,620
MF Coal to Offsites, t/d	4,630
Electricity, MW	-
Output, t/d (bbl/d)	
Fuel Gas	784
Propane (LPG)	69
Butane (LPG)	11
Gasoline	1,055 (9340)
Diesel Oil	214 (1785)
Fuel Oil	63 (483)
Alcohols	133 (1047)
Sulfur	120
Liquid Product* Yield (based on coal to gasifier)	
bbl/t as-received coal	0.65
bbl/t MF coal	1.0
bbl/t MAF coal	1.26
Liquid Product* Yield (based on coal to plant)	
bbl/t as-received coal	0.46
bbl/t MF coal	0.71
bbl/t MAF coal	0.91
Thermal Efficiency, %	
Overall	36.4
Liquid Products	18.9

* Liquid Products include gasoline, diesel oil and fuel oil

Table 4-5 Comparison of Product Yields (Base Case vs Saskoil Case)
[Basis: 1000 t of MAF coal to gasifier*]

Products**	Base Case (tonne)	Saskoil Case (tonne)
SNG**	248.0	85.4
Propane	6.4	7.5
Butane	0.9	1.0
Gasoline	110.8	114.9
Diesel Oil	20.0	23.3
Fuel Oil	5.8	6.8
Alcohols	16.7	15.3

* Lurgi dry-ash gasifier in the base case, Koppers-Totzek gasifier in the Saskoil case

** SNG Product quality is not the same. In the base case, 90% CH_4 and 3.8% H_2. In Saskoil case, 50% CH_4 and 48% H_2. All other products have the same quality in both cases.

is summarized in Table 4-6. The plant is self-sufficient in terms of utilities requirement. About 48,696 t/d (8,936 gpm) of make-up water is required from an outside source. As discussed earlier in Section 2.2 (see coal-to-methanol plant material balance), if the cooling air is employed for process streams wherever possible, the make-up water requirements may decrease by approximately 50%. The electrical power consumption is about 92 MW which is generated in the offsites. This power consumption is based on the plant configuration where steam turbines are used for power intensive operations such as raw gas compression, air compression and fuel gas compression.

The overall thermal efficiency of the conceptual commercial plant operated with Willow Bunch lignite is 36.4%. The input energy includes coal to gasifiers for synthesis gas production and to offsite boilers to generate steam and power for the plant. The output energy includes product and byproduct energy.

The liquid product (i.e. gasoline, diesel and fuel oil) efficiency of the plant is 18.9%. If propane and butane (LPG) are also included, the liquid product efficiency increases to 20%.

Table 4-6 Summary of Utilities Balance of a Self-Sufficient SASOL-Type Plant via Fischer-Tropsch Synthesis
[Basis: 25,000 t/d "as-received" Willow Bunch lignite]

UTILITY / SECTIONS	Steam* 10.8 MPa t/d	Steam* 3.6 MPa t/d	Steam 3.6 MPa t/d	Steam 0.25 MPa t/d	Condensate Water t/d	Boiler** Feed Water t/d	Cooling Water t/d	Raw*** Water t/d	Electricity kWh/h
Lignite Preparation				312			5,280		18,910
Lignite Gasification						19,200	30,720		22,770
Gas Treatment		11,280	8,496				513,600		9,980
Fischer-Tropsch Synthesis	3,792		3,048	3,768		12,000	75,624		9,400
Methanation	1,272		504	1,440		2,160	7,608		650
Oxygen Plant		11,808					240,000		720
Emission Control									220
Effluent Control	2,544			5,448					1,150
Product Blending									160
Utilities	21,264		4,416	1,680		27,456	1,008,000		28,200
Offsites									260
TOTAL (CONSUMPTION)	28,872	23,088	16,464	12,648	-	60,816	1,880,032	-	92,200
Lignite Preparation					312				
Lignite Gasification			12,912	5,760					
Gas Treatment					14,688				
Fischer-Tropsch Synthesis				5,448	10,608				
Methanation			1,296		3,696				
Oxygen Plant					11,808				
Emission Control									
Effluent Control		2,544							
Product Blending	28,872								
Utilities		20,544	2,256	1,440	6,816		1,839,576	48,696	
Offsites									92,200
TOTAL (PRODUCTION)	28,872	23,088	16,464	12,648	47,928	-	1,839,576	48,696	92,200

* Superheated steam
** Includes blow down water
*** Boiler feed and cooling tower make-up water

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4.4 PRODUCT QUALITY

4.4.1 **Gasoline**

The composition of blended gasoline product from the Fischer-Tropsch synthesis plant is given below:

	wt%
Mixed Butanes	3.0
Light Hydrocarbons	3.3
Alkylate	4.4
Light Polymer Gasoline	17.8
Hydrogenated Heavy Polymer Gasoline	16.6
C_5/C_6 Isomerate	28.2
Reformate	26.7
	100.0

Estimated properties and a comparison with unleaded gasoline specifications are presented in Table 4-7. This gasoline meets all specifications. Compared to typical present-day gasoline, it is slightly lower in gravity and contains fewer aromatics.

While the olefin content of 20 vol% is considered tolerable, the quality of the F-T gasoline is poor compared with the MTG gasoline which contains 11 vol% olefins (see Table 3-5).

4.4.2 **Diesel Oil**

Estimated properties and a comparison with diesel fuel specifications are shown in Table 4-8. This product could be marketed as a premium diesel oil, Grade No. 1-D, for engine service requiring frequent speed and load changes. The estimated cloud point of

Table 4-7 Comparison of Principal Unleaded Gasoline Specifications with Estimated Fischer-Tropsch Gasoline Properties

	Estimated MTG Unleaded Gasoline Properties	Specification
Gravity, °API	67.2	
Octane Numbers		
Research	91	
Motor	83	82 min.
(Research + Motor)/2	87	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation, °C		
IBP	30	
10%	42.2	70 max.
30%	74.4	
50%	85.6	76.7/121 min./max.
70%	120.6	
90%	168.3	190 max.
EP	215.6	225 max.
V/L Ratio (=20), °C	52.8	@ 60 max.
Sulfur, wt.%	Nil	0.10 max.
Composition, vol.%		
Paraffins	60	
Olefins	20	20 max. (target)
Naphthenes	3	
Aromatics	17	
Molecular Weight	93	

Source: DOE Report FE-2447-9, Jan. 1978

Table 4-8 Comparison of Principal Diesel Oil Specifications with Estimated Fischer-Tropsch Diesel Oil Properties

	DIESEL OIL ASTM D975* Grade No. 1-D	Grade No. 2-D	Other	Estimated F-T Diesel Oil Properties
Gravity, °API	-	-		56
Cetane Number, min. (D613 or D976)	40	40	45**	50
Volatility				
90% min./max., °F (D86)	-/550	540/640 ##		520
Flash Point, °F (D93)	100	125		125
Cloud Point, °F (D2500)	***	***		10 - 25
Viscosity, Kin. @ 100°F min./max. (D445)	1.4/2.5	2.0/4.3		2
Sulfur, wt% max. (D129)	0.50	0.50	#	None
Corrosion, max. (D130)	No. 3	No. 3		No. 1
Carbon Residue on 10% Resid. max. (D524)	0.15	0.35		Trace

* 1976 Annual Book of ASTM Standards
** Various engine builders' requirements
*** Varies with marketing region; maps included in ASTM D975
# Varies in individual states, some as low as 0.3%
## 540/680 90% min./max. proposed but currently not accepted by industry

Source: DOE Report FE-2447-9, Jan. 1978

4.4.2 Diesel Oil (cont'd)

this oil is however high. This could preclude its marketing in cold weather regions.

4.4.3 Heavy Fuel Oil

Estimated properties and a comparison with gas turbine fuel oil specifications are presented in Table 4-9. This product is suitable for use as a premium gas turbine fuel particularly because of non-existence of sulfur and metal content.

4.4.4 Alcohols

The alcohol product is essentially free of acids, aldehydes, ketones and water. The estimated composition of the mixed alcohol product is as follows:

	wt%
ethanol	59
i - propanol	14
n - propanol	14
2 - butanol	4
i - butanol	1
n - butanol	5
2 - pentanol	1
n - pentanol	1
C_6^+	1
	100

Table 4-9 Comparison of Principal Fuel Oil and Gas Turbine Fuel Oil Specifications with Estimated Fischer-Tropsch Heavy Fuel Oil Properties

	Fuel Oil No. 5 (Light) ASTM D396 *	Gas Turbine F.D. No. 3 - GT ASTM D2880*	Estimated F-T Heavy Fuel Oil Properties
Gravity, °API	-	-	41
Volatility			
Flash Point, °F min. (D93)	130 or Legal	130 or Legal	150
Distillation	-	-	-
Ash, wt% max. (D482)	0.10	0.03	Trace
Viscosity, SSU at 100°F, min. (D445)	150	45	Satisfactory
Sulfur, wt% (D129)	Legal	-	None
Metals, ppm (wt.) max.			
Vanadium	-	2	0
Sodium + Potassium	-	5	Trace
Calcium	-	10	0
Lead	-	5	0

* 1976 Annual Book of ASTM Standards

Source: DOE Report FE-2447-9, Jan. 1978

4.4.5 Propane and Butane LPG

The estimated properties of propane and butane LPG are presented in Table 4-10 along with ASTM specifications. The comparison indicates that both of these products conform to commercial specifications.

4.4.6 Fuel Gas

The estimated compostion of fuel gas is given below:

	vol%
Hydrogen	50.0
Methane	48.5
Carbon Dioxide	0.1
Inerts (N_2 + Ar)	1.3
Others	0.1
	100.0

As indicated previously, syngas produced in the Koppers-Totzek gasifier does not contain methane. The main source of methane in the fuel gas is from the F-T synthesis. Although the product fuel gas has high calorific value (GCV = 62460 kJ/kg), it does not conform to SNG specification (90 vol% CH_4, 4 vol% H_2). This product gas could be marketed to utility companies. Alternatively, it could be blended with pure methane to conform to SNG specifications or hydrogen could be separated and sold as byproduct.

Table 4-10 Comparison of Principal LPG Specifications with Estimated Fischer-Tropsch LPG Properties

	Commercial Propane ASTM D1835*	Est. F-T C_3 LPG Properties	Commercial Butane ASTM D1835*	Est. F-T C_4 LPG Properties
Vapor Pressure at 100°F, max. psig (D1267)	208	179	70	37
Volatile Residue				
Butane and Heavier, max. vol% (D2163)	2.5	1	-	-
Pentane and Heavier, max. vol% (D2163)	-	-	2.0	1
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

* 1976 Annual Book of ASTM Standards

Source: DOE Report, FE-2447-9, Jan. 1978

5.0 ENVIRONMENTAL ASPECTS

5.0 ENVIRONMENTAL ASPECTS

Various solid, liquid and gaseous wastes will be generated in the indirect lignite liquefaction schemes presented earlier. The environmental discharge of these wastes will have to be in compliance with the regulatory environment control acts of the province of Saskatchewan.

At present, there are five environment control acts covering air pollution, water resources management, water rights, ground water conservation and public health. A summary of provisions and requirements under these acts is presented in Appendix D. The environmental aspects pertaining to the indirect liquefaction of Willow Bunch lignite are outlined in this chapter. Detailed environmental assessment or environmental impact statement is beyond the scope of the study.

5.1 KOPPER-TOTZEK GASIFICATION PLANT

- Solid Wastes

Large amounts of solid waste will be generated in the lignite preparation and gasification area. This waste will mainly consist of slag from gasifier bottom (1060 t/d), ash from offsite boilers (1200 t/d) and fly ash from the electrostatic precipitators. The principal constituents of the solid waste would be the mineral composition of Willow Bunch lignite. The gasifier slag is granulated and extracted in a water bath before its disposal to mine site. This will reduce the possibility of leaching of trace metals from the slag into ground or surface waters. Ash from offsite boilers and fly ash will be disposed as a slurry cake through the open ash lagoon system. The system will require a water make-up supply due to evaporation losses from the lagoon and water trapped in the lagooned ash. Since good quality water is not required for this service, cooling water blowdown will be used.

The dissolved solid content of the water will be high due to evaporation losses and ash leaching. Hence the lagoon will have a purge stream for the disposal of some of the supernatant matter. The ash disposal in this manner lowers the leaching of trace metals into ground or surface waters and therefore will be environmentally acceptable.

- Liquid Wastes

Liquid waste will consist of sludge from the thickener of the raw gas wash water system. Waste sludges will also be generated during the demineralization of raw water and boiler feed water treatment. Another major contaminated stream will be sour water generated by the wet scrubbers where the raw gases produced by the gasifier are cleaned. The compositions of thickener sludge and waste water as given by Krupp-Koppers are presented in Table 5-1.

The environmental assessment of waste water stream from a commercial Kopper-Totzek gasification plant* has been reported in "Environmental Assessment: Source Test and Evaluation Report" prepared for U.S. Environmental Protection agency (C.A. Zee et al, EPA report 600/7-81-009, Jan. 1981). The process configuration of this plant to produce purified synthesis gas is similar to the process configuration described in Section 2.1. Based on the results of the above mentioned study, the proposed conceptual indirect liquefaction plant operating with Willow Bunch lignite would meet the environmental regulations for liquid wastes.

- Gaseous Emissions

The proposed configuration of the indirect liquefaction plant includes an emission control section consisting of Claus sulfur plant and SCOT tail gas treatment unit. The various gaseous emission streams from the entire plant are combined and treated in this unit which is designed to control the SO_2 emission within the range of 100 to 140 ppm. This level of SO_2 discharge to atmosphere is environmentally acceptable under Saskatchewan's air quality standards (see Table D-1, Appendix D).

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* 1000 t/d ammonia plant operating at Modderfontein, South Africa since 1974. Feed coal contains 20.7 wt% ash, 0.5 wt% sulfur and 1.6 wt% nitrogen on MF basis.

Table 5-1 Waste Streams in the Gasification of Willow Bunch Lignite

Solid Waste		
- Slag		
Origin	Gasifier	
Quantity	1060 t/d	
Size	0 to 15 mm	
Temperature	50 to 70°C	
Bulk Density	About 1.2 t/m	
Moisture	Dripping Wet	
Composition	Similar to the Ash Analysis of Willow Bunch Lignite	
- Ash		
Origin	Lignite Preparation Steam and Power Generation	
Quantity	1200 t/d	
Composition	Similar to the ash analysis of Willow Bunch Lignite	
Liquid Waste		
- Slurry		
Origin	Thickener of Wash Water System	
Quantity	450 m^3/h	
Composition	Water :	90 wt%
	Fly ash (solids) :	10 wt%
	HCN :	1 mg/L
Temperature	50°C	
- Waste Water (Water Treatment)		
Origin	Demineralization	
Quanitity	80 t/d	
pH value	6.5 to 8.5	
Salt Content	Ca^{++}	500 mg/L
	Mg^{++}	90 mg/L
	Na^{+}	2,840 mg/L
	NO$_3^-$	25 mg/L
	Cl$^-$	400 mg/L
	SO$_4^{--}$	5,860 mg/L
	S$_1$O$_2$	630 mg/L
	HCO$_3^-$	710 mg/L
	Total	11,055 mg/L

Table 5-1 Waste Streams in the Gasification of Willow Bunch Lignite (cont'd)

Gaseous Waste	
- Flue Gas	
Origin	Claus Incinerator
Quantity	2,040,000 m^3/h
Analysis, dry gas	CO_2 16 vol% N_2 84 vol% SO_2 130 ppm Dust 100 mg/m^3

5.2 METHANOL SYNTHESIS PLANT

Methanol synthesis plants have been operating commercially and are designed to meet environmental regulations. The waste generated in the methanol synthesis unit is solid waste containing synthesis catalyst and spent zinc oxide. These wastes can be disposed to the mine with other solid wastes.

5.3 MOBIL-MTG PLANT

Emissions from the methanol-to-gasoline conversion unit will include MTG catalyst regeneration off-gas, spent dimethyl ether and MTG catalyst. Catalyst regeneration off-gas which would contain nitrogen, carbon dioxide, oxygen, water vapor, and some hydrocarbons (flushed out of the reactor at the regeneration cycle) will be incinerated. Spent DME and MTG catalyst will be disposed with the other solid wastes.

5.4 FISCHER-TROPSCH SYNTHESIS PLANT

The environmental aspect of the Fischer-Tropsch conversion technology has been discussed in the ERDA Report No. 114, January 1977, prepared by the Ralph M. Parsons Company.

No aqueous effluent standards specifically addressed to coal conversion plants have been issued by the state legislatures or by the Federal Government. Standards that are somewhat related to Fischer-Tropsch process would be the standards issued for petroleum refining. The results of Parson's Report (based on bituminous coal containing 1.1 wt% nitrogen, 3.4 wt% sulfur on MF basis) indicate that the aqueous effluents estimated for Fischer-Tropsch plant are either the same or lower than the standard parameters outlined by the U.S. Environmental Protection agency for petroleum refineries (EPA Report EPA-440/1-74-014a, Washington D.C., April 1974).

5.5 OCCUPATIONAL SAFETY

Of particular interest in coal conversion projects is the possible formation of carcinogenic compounds in the course of coal hydrogenation and pyrolysis. These compounds are usually found in coal tars, and are not expected to be formed under the operating conditions of the entrained-flow gasifier selected for this study.

Carcinogenic activity of higher boiling fractions obtained from petroleum refining has also been observed on laboratory animals. Similar fractions are obtained on distillation of the liquid hydrocarbons produced by the Fischer-Tropsch reactor, and Fischer-Tropsch oils boiling above 250°C were found carcinogenic in mice*. A corresponding activity in man, however, has not been shown; cancer frequency in oil refinery workers is the same as for other industrial occupations. The amount of known carcinogens, such as benzo(a)pyrene, in high boiling petroleum fractions is much lower than in coal tar; the products are also maintained in close systems, so that workers have very little contact with the product. It is felt that similar conditions will exist in Fischer-Tropsch operations, thereby minimizing environmental risk.

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* W.C. Hueper, "Experimental Carcinogenic Studies on Hydrogenated Coal Oils. II. Fischer-Tropsch Oils", Industrial Medicine and Surgery, Vol. 25, pp. 459-62, 1956.

6.0 ECONOMIC ASSESSMENT OF INDIRECT COAL LIQUEFACTION PROCESSES

6.0 ECONOMIC ASSESSMENT OF INDIRECT COAL LIQUEFACTION PROCESSES

Preliminary capital cost estimate, operating, maintenance and product cost estimate of three indirect coal liquefaction alternatives have been performed for a self-sufficient, integrated plant processing 25,000 t/d of Willow Bunch lignite. These cost have been developed in accordance with the criteria described earlier in the text and following the procedure developed by SNC and approved by the Study Committee.

All three indirect coal liquefaction process alternatives employ Koppers-Totzek gasifier configuration to convert coal to synthesis gas for the production of:

- pure methanol (99.9 wt% purity) via low-pressure methanol synthesis developed by Imperial Chemical Industries (ICI process).
- motor gasoline via Mobil's Methanol-to-Gasoline (MTG) route using ICI process for methanol synthesis.
- production of gasoline and hydrocarbons via SASOL-type Fischer-Tropsch process employing Synthol fluidized-bed reactor.

For each of the three process alternatives, product costs have been calculated on "Energy Value" and "Product Value" bases. Product costs on"Energy Value" basis is calculated considering heating value only whereas the other also considers the market value (quality) of the product. The "Product Value" approach is also known as ESCOE* methodology.

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* K.A. Rogers and R.F. Hill, the Engineering Society Commission on Energy Inc., "Coal Conversion Comparison", DOE Report No. FE-2468-51, July 1979

6.1 FINANCIAL ASSUMPTIONS AND ECONOMIC CRITERIA

Some of the basic financial assumptions and criteria for evaluation of coal conversion processes are summarized in Table 6-1. Other data and economic criteria received from the process developers and the Study Committee are presented later in the text.

All three indirect coal liquefaction process alternatives are evaluated as a self-sufficient, integrated, grass-roots facilities designed to process 25,000 t/d of as-received Willow Bunch lignite supplied to the plant. The plant is assumed to be located within 16 km (10 miles) of the coal mines in Willow Bunch area, Saskatchewan. Coal, air and raw water are the major process inputs to the plant. All other utility requirements such as steam, fuel gas and electricity are generated on site.

The feed coal to the plant is assumed to be sized to minus 5-cm and having typical (average) ultimate and proximate analyses as shown in Appendix A.

All costs are calculated in "constant" Canadian mid-1982 dollars and no forward escalation beyond mid-1982 is included. Cost estimates and data received from process developers or licensors for a different plant location, capacity and conditions have been converted and adjusted according to standard industry index.

For the plant, a 4-year construction period, 30-year operating life, straight line depreciation and zero salvage value are assumed. The capital investment schedule during construction is based on a typical S-type curve (see Appendix E-1) and the full design-capacity level is expected to be achieved within 3 years of the start-up date (see Appendix E-2).

Table 6-1 Financial Assumptions and Economic Criteria

Coal feed: 25,000 tonnes per day to plant.

Constant Canadian dollars, mid-1982.

Plant location: within 16 km (10 miles) of coal mines

4-years construction period.

Capital investment schedule (S-type curve)

Year	%
- 4	9
- 3	25
- 2	36
- 1	30

Investment tax credit for cost of borrowing during construction claimed during year of occurence.

Plant life: 30 years

Straight-line depreciation method

Zero salvage value

Process contingency: 5%

Project and estimating contingency: 10%

Plant start-up efficiencies

Year	% Design Capacity
1	50
2	90
3	100

Plant service factor: 0.85 (310 operating days per year)

Cost of capital financing = 10% per annum

Land cost: $990/hectare ($400/acre)

Working capital: 3% of installed-plant cost

6.1 FINANCIAL ASSUMPTIONS AND ECONOMIC CRITERIA (cont'd)

Cost of capital is assumed to be 10% per annum. For 100% debt financing, the value is equivalent to debt cost* in "real" terms. For 100% equity financing, it may be considered as return on equity before taxes.

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* For calculation in "constant" dollars, the debt cost is expressed in real terms i.e., prevailing rates minus inflation.

6.2 CAPITAL COST ESTIMATE AND TOTAL INVESTMENT

Capital cost estimates and investments data available from the process developers/licensers and open literature for a different plant location have been adjusted to the designated Saskatchewan area utilizing the criteria and procedures approved by the Study Committee which is described below.

- Equipment items are assumed to be 25% Canadian content and 75% U.S. content. Items purchased in U.S. are not available in Canada, otherwise, a duty of 12.8% will be levied (Customs Canada, 27 Jan. 1982).

- All bulk materials (pipings, fittings, etc.,) are considered as 100% Canadian content, comprising 45% labour and 55% material components.

- The U.S. material and equipment content (75% of the total) is converted to Saskatchewan base by multiplying by a factor of 1.19 composed strictly of currency exchange rate (Royal Bank 27-01-82).

- The Canadian material and equipment content (25% of the total) is obtained by multiplying the cost given for a U.S. location by a shop-labour price differential factor of 1.1592 (see Appendix E-3). Including the exchange rate of 1.19, the "material plus currency differential factor" is 1.379.

- Construction manpower/labor costs for a reference location are converted to Saskatchewan location by multiplying by a productivity adjustment factor of 1.42 (see Appendix E-4).

- Capital cost estimates are adjusted for inflation to the base year (mid-1982) using the Chemical and Mineral Plant Price Index given in Appendix E-5.

6.2.1 Coal-Based Plant for Production of Pure Methanol

Following the above procedure, the capital cost estimate of a methanol plant (Koppers-Totzek gasifier configuration with ICI low-pressure methanol synthesis) with breakdown by individual processing area/units is given in Table 6-2. All costs except those for the SCOT tail-gas processing unit, flue gas desulfurization, ancillary and offsite facilities have been based on Krupp-Koppers data*. Cost for the coal preparation section supplied by Krupp- Koppers has been increased by 10% to allow for coal receiving, storage and reclaiming facilities. Krupp-Koppers data in U.S. dollars (January 1982) has been converted to Canadian dollars (mid-1982) and adjusted for the designated Saskatchewan location.

The SCOT tail-gas processing unit not included by Krupp-Koppers has been added in order to meet the environmental regulations for sulfur emission. The sulfur content in the vent gases are consequently reduced from 900 ppm to less than 130 ppm. Capital cost of the SCOT unit is based on the cost of a similar unit estimated by Oak Ridge National Laboratory**. The estimate in mid-1979 U.S. dollars provided in ORNL report has been adjusted for capacity difference, exchange rate, material and labor rate differential following the guidelines described earlier.

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* H. Staege, Krupp-Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1913, June 1982.

** R.M. Wham et al, ORNL Report No. 5644, February 1981.

Table 6-2 Capital Cost Estimate of a Coal Based Plant for the Production of Pure Methanol, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Section	Process Area	Equipment Items, Engineering and License Fees	Civil Works, Bulk Materials and Labor	Installed Cost
100	Coal Preparation *	84.24	112.12	196.36
200	Coal Gasification	149.40	78.57	227.97
300	Raw Gas Compression	53.98	25.49	79.47
	Acid Gas Removal (Rectisol)	76.58	40.34	116.92
	Shift Conversion	32.64	16.98	49.62
400	Methanol Plant (including MeOH Factionation)	72.82	38.22	111.04
500	Oxygen Plant	104.20	57.33	161.53
600	Claus Sulfur Recovering Unit	7.53	4.23	11.76
	SCOT Tail-Gas Processing Unit	3.90	2.32	6.22
700	Effluent Control (Wash Water Treatment)	10.04	10.62	20.66
800	Product Storage and Blending	**	**	**

* Cost supplied by Krupp-Koppers has been increased by 10% to include coal receiving, storage, and reclaiming facilities.

** Included in the respective processing areas

Table 6-2 Capital Cost Estimate of a Coal Based Plant for the Production of Pure Methanol, in Millions of Canadian Dollars (mid-1982) (cont'd)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Section	Process Area	Equipment Items, Engineering and License Fees	Civil Works, Bulk Materials and Labor	Installed Cost
900/1000	Steam and Power Generation	99.18	50.95	150.13
	Energy Distribution	15.06	6.37	21.43
	Fuel Gas Desulfurization	46.68	64.97	111.65
	Cooling Tower System	15.06	8.48	23.54
	Feed Water System	17.58	8.48	26.06
	Ancillary Units*	24.58	14.63	39.21
	Offsite Facilities	80.23	47.73	127.96
1100	Product Upgrading and Refining (Methanol Fracturation)	**	**	**
	SUB-TOTAL	893.70	587.83	1,481.53
	Adjustment Factor for US and Canadian Content (0.25 x 1.1592 +0.75 =1.0398)	35.57	-	35.57
	TOTAL	929.27	587.83	1,517.10

* Includes instrumentation air, biological waste-water treatment, surface and storm water treatment, and water retention ponds.

** Included in the methanol plant (Section 400)

Sources: H. Staege, Krupp-Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982.
R.M. Wham et al., ORNL Report No. 5644, February 1981.
W.P. Syrett, NCB report to SNC Inc., NCB Ref. No. DR-496/2, 7/3, January 23, 1982
SNC in-house information, original data supplied by Shell Development Company, Process Licensing Division, Houston, Texas.

6.2.1 Coal-Based Plant for Production of Pure Methanol (cont'd)

Other miscellaneous items not included by Krupp-Koppers in their capital cost estimate have been added under ancillary and offsite facilities. Ancillary units include instrument air supply, biological waste-water treatment, surface and storm water treatment and waste-water retention ponds. Offsites include product handling and shipping facilities, relief system and flare, incinerator, fire protection equipments and shed, buildings, roads, parking lots, fencing, etc. Ancillary and offsite facilities have been estimated from ORNL report.

Finally, the cost of flue gas desulfurization unit is also added. The reference data in this case is supplied by the National Coal Board* to SNC Inc. for Willow Bunch lignite.

The total capital investment for the methanol plant is given in Table 6-3. The installed-plant includes all the equipment items, bulk materials, labor, engineering and support, process contingency, licenses and prepaid royalties and other construction costs. A project and estimation contingency of 10% has been added to cover uncertainties in cost estimation, equipment damage during construction, delay in equipment delivery and project schedule.

Initial catalysts, chemicals and operating supplies cover the cost of "initial charge" of these materials to the plant. These costs have been assumed to be equivalent to a 1-month supply of catalysts, chemicals and other consumables (except water) based on full-plant capacity.

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* W.P. Syrett, NCB report to SNC Inc., DR 496/2, 7/3, January 23, 1982

6.2.1 **Coal-Based Plant for Production of Pure Methanol** (cont'd)

Start-up and other costs taken as 3% of total plant installed cost (TPIC) cover operating expenses during commissioning and start-up period. These costs also include additional capital for modifications required to bring the plant to full capacity, associated process and environmental studies, contractors' fees, owners' engineering offices and cost for obtaining construction permits.

Working capital is assumed as 3% of TPIC. The size of land has been estimated as 80 hectare (200 acres*) from the plot plan supplied by Krupp-Koppers GmbH. The land rate of $900/hectare ($400/acre) is assumed.

Interest during construction is based on capital spending schedules given in Table 6-3 and Appendix E-2. Investment credits equal to 48% of interest during construction is claimed in the year of occurrence.

The total investment of a 25,000-t/d coal-based self-sufficient, commercial methanol plant (Koppers-Totzek gasifier configuration for synthesis gas production and ICI low-pressure methanol synthesis) including all the processing units, utilities and ancillary facilities is Canadian $1925.74 million.

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* 1 acre = 4047m^2

Table 6-3 Breakdown of Total Capital Investment of a Coal Based Plant for Production of Pure Methanol, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of "as-received" Willow Bunch lignite]

Installed-Plant Cost (including process contingencies)	1,517.10
Project and Estimation Contingencies	151.71
Total Plant Installation Cost (TPIC)	1,668.81
Initial Catalysts, Chemicals and Operating Supplies	1.40
Start-up and Other Costs	50.06
Working Capital	50.06
Land	0.08
Interest During Construction	155.33
Total Capital Investment, "C"	1,925.74

Sources:

H. Staege, Krupp-Koppers report to SNC Inc., VK-1/Mn-Ko-1983, June 1982

B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979.

6.2.1 **Coal-Based Plant for Production of Pure Methanol** (cont'd)

The total investment, however, does not include:

- The cost of coal mine, mine development and transportation facilities from the mine to the plant.
- Railways, roads and other transportation facilities outside the plant site.
- Product distribution pipelines.
- Water supply and pipelines outside the plant site.
- Housing, community facilities and associated infrastructure.

6.2.2 **Coal-Based Plant for Production of Gasoline via Mobil Methanol-to-Gasoline (MTG) Process**

The capital cost estimate for a coal-based plant for the production of gasoline via Mobil Methanol-to-Gasoline (MTG) process is given in Table 6-4. The plant includes Koppers-Totzek gasifier configuration for production of synthesis gas and ICI low-pressure process for methanol synthesis.

The data, basic sources and procedures used for capital cost estimate of this plant are similar to that described earlier for a methanol plant (Subsection 6.2.1) except that:

- cost of methanol fractionation unit has been deleted since Mobil MTG process does not require pure methanol.

- cost of waste-water treatment unit (Can. $19.93 million) has been added to effluent control (Section 700) to process additional 670 U.S. gpm (42.3 L/s) of sour water produced in MTG unit. The reference data (Report No. ORNL-5664, Feb. 1981) in mid-1979 U.S. dollars has been adjusted for capacity, exchange rate, material and labour rate differences and adjusted for Saskatchewan location.

Table 6-4 Capital Cost Estimate of a Coal-Based Plant for the Production of Gasoline (Mobil-MTG), in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Section	Process Area	Equipment Items, Engineering and License Fees	Civil Works, Bulk Materials and Labor	Installed Cost
100	Coal Preparation *	84.24	112.12	196.36
200	Coal Gasification	149.40	78.57	227.97
300	Raw Gas Compression	53.98	25.49	79.47
	Acid Gas Removal (Rectisol)	76.58	40.34	116.92
	Shift Conversion	32.64	16.98	49.62
400	Methanol Plant	62.68	31.16	93.84
500	Oxygen Plant	104.20	57.33	161.53
600	Claus Sulfur Recovering Unit	7.53	4.23	11.76
	SCOT Tail-Gas Processing Unit	3.90	2.32	6.22
700	Effluent Control (includes Wash-Water and Waste Water Treatment)	22.62	17.97	40.59
800	Gasoline Blending **	2.97	1.75	4.72

* Cost supplied by Krupp-Koppers has been increased by 10% to include coal receiving, storage, and reclaiming facilities.

** Product storage cost is included in respective processing section/areas.

Table 6-4 Capital Cost Estimate of a Coal-Based Plant for the Production of Gasoline (Mobil-MTG), in Millions of Canadian Dollars (mid-1982) (cont'd)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Section	Process Area	Equipment Items, Engineering and License Fees	Civil Works, Bulk Materials and Labor	Installed Cost
900/1000	Steam and Power Generation	99.78	51.35	151.13
	Energy Distribution	15.06	6.37	21.43
	Fuel Gas Desulfurization	46.68	64.97	111.65
	Cooling Tower System	14.32	8.06	22.38
	Feed Water System	17.58	8.48	26.06
	Ancillary Units*	27.04	16.09	43.13
	Offsite Facilities	80.23	47.73	127.96
1100	Methanol Conversion (MTG)	119.51	69.80	189.31
	Fractionation	5.87	3.43	9.30
	HF Alkylation	7.00	4.11	11.11
	SUB-TOTAL	1,033.61	668.65	1,702.26
	Adjustment Factor for US and Canadian Content (0.25 x 1.1592 +0.75 =1.0398)	41.14	-	41.14
	TOTAL	1,074.75	668.65	1,743.40

* Includes instrumentation air, biological waste-water treatment, surface and storm water treatment, and water retention ponds.

Sources: H. Staege, Krupp-Koppers GmbH report to SNC Inc. VK-1/Mn-Ko-1813, June 1982.
R.M. Wham et.al., ORNL Report No. 5644, February 1981
W.P. Syrett, NCB report to SNC Inc., NCB Ref. No. DR-496/2, 7/3, January 23, 1982
SNC in-house information, original data supplied by Shell Development Company, Process Licensing Division, Houston, Texas

6.2.2 Coal Based Plant for Production of Gasoline via Mobil Methanol-to-Gasoline (MTG) Process (cont'd)

- cost of gasoline blending unit (Can. $4.72 million) has been added. (reference data: Report No. ORNL-5664, Feb. 1981)

- cost of steam and power generation unit and cooling water system has been pro-rated to account for capacity differences.

- cost of Mobil-MTG processing unit for methanol conversion, fractionation and HF alkylation has been added. The reference data in October-1977 U.S. dollars are taken from Mobil study (DOE Report No. FE-2447-13, August 1978). Process contingencies of 5% and prepaid royalties of 1% (not included in Mobil's data) have been added to present the data on a consistent basis.

- cost of ancillary units and offsite facilities has been increased by 10% to account for additional processing required by Mobil-MTG process.

The total capital investment of a self-sufficient commercial plant processing 25,000 t/d of as-received Willow Bunch lignite is Canadian $2212.76 million as shown in Table 6-5. The installed cost of the plant includes process equipment items, utilities, ancillary and offsites facilities, process contingencies, licenses and prepaid royalties. Other items are project and estimating contingencies of 10%, cost of "initial" charge of catalysts and chemicals, start-up cost, working capital, land, and interest during construction.

Table 6-5 Breakdown of Total Capital Investment of a Coal-Based Plant for Production of Gasoline via Mobil-MTG Process, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Installed-Plant Cost (including process contingencies)	1,743.4
Project and Estimation Contingencies	174.3
Total Plant Installation Cost (TPIC)	1,917.7
Initial Catalysts, Chemicals and Operating Supplies	1.5
Start-up and Other Costs	57.5
Working Capital	57.5
Land	0.08
Interest During Construction	178.48
Total Capital Investment, "C"	2,212.76

Sources:

H. Staege, Krupp-Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982
M. Schreiner, Mobil Research and Development Corporation, DOE Report No. FE-2447-13, August 1978.
B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979.

6.2.3 Coal-Based Plant for Production of Gasoline and Other Hydrocarbons via Fischer-Tropsch Synthesis

The capital cost estimate of SASOL-type coal liquefacation plant for production of gasoline and other hydrocarbons via Fischer-Tropsch synthesis (Synthol reactor) is given in Table 6-6.

Items marked by single asterisk (*) are based on Krupp-Koppers data for production of synthesis gas from 25,000 t/d of Willow Bunch lignite. The cost of coal preparation section has been increased by 10% to account for coal receiving, storage and reclaiming facilities not included in the Krupp-Koppers estimate.

Items marked by two asterisks (**) are taken from ORNL study (Report No. ORNL-5664, Feb. 1981). The cost of the waste-water system has been added to treat an additional 740 U.S. gpm (46.7 L/s) of sour water from Fischer-Tropsch synthesis unit (Section 200A). Ancillary units and offsite facilities not included by Krupp-Koppers have been added. Ancillary units include instrument air supply, biological waste-water treatment, surface and storm water treatment, and waste-water retention ponds. Offsite facilities are the same as the items listed for methanol plant plus Fischer-Tropsch catalyst preparation unit.

Costs of all other processing areas, i.e., Fischer-Tropsch synthesis (Section 200), methanation, CO_2 removal, and SNG drying units (Section 400) are based on Mobil study (DOE Report FE-2447-13, Aug. 1978).

The total capital investment for a self-sufficient, integrated SASOL type plant with Kopper-Totzek gasifier configuration and Fischer-Tropsch synthesis (Synthol reactor) is given in

Table 6-6 Capital Cost Estimate of a SASOL-Type Coal Liquefaction Plant via Fischer-Tropsch Synthesis (Synthol Reactor), in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Section	Process Area	Equipment Items, Engineering and License Fees	Civil Works, Bulk Materials and Labor	Installed Cost
100	Coal Preparation *	85.62	87.66	173.28
200	Coal Gasification *	155.67	63.09	218.76
200A	Synthol Reactor (F-T Synthesis)	100.26	58.89	159.15
	F-T Product Fractionation	21.24	12.48	33.72
	F-T Product Hydrotreating	10.98	6.46	17.44
	Hydrotreater Product Fractionation	6.38	3.75	10.13
	Catalytic Reforming	12.26	7.21	19.47
	C_5/C_6 Isomerization	8.94	5.24	14.18
	Catalytic Polymerization	6.89	4.04	10.93
	HF Alkylation	3.69	19.09	22.78
	Polymer Gasoline Hydrogenation	3.99	2.34	6.33
	Light-Ends Recovery	3.60	2.11	5.71
	H_2 Purification	3.44	2.02	5.46
	Alcohol Recovery	28.25	16.63	44.88
	Hydrocarbon Recovery (including refrigeration)	55.89	27.86	83.75
	Hydrogen Recovery	7.10	4.17	11.27
300	Raw Gas Compression *	56.49	23.09	79.58
	Acid Gas Removal *	79.09	35.53	114.62
	Shift Conversion *	33.90	15.99	49.89
400	Methanation	11.75	6.89	18.64
	CO_2 Removal	8.45	4.97	13.42
	SNG Drying & Compression	3.20	1,89	5.09
500	Oxygen Plant *	116.71	49.69	156.4
600	Sulfur Plant *	8.79	3.55	12.34
	SCOT Unit *	3.92	2.37	6.29

* Items designated by an asterisk have been calculated using Krupp-Koppers data.

Table 6-6 Capital Cost Estimate of a SASOL-Type Coal Liquefaction Plant via Fischer-Tropsch Synthesis (Synthol Reactor), in Millions of Canadian Dollars (mid-1982) (cont'd)

[Capacity: 25,000 t/d of Willow Bunch lignite]

Section	Process Area	Equipment Items, Engineering and License Fees	Civil Works, Bulk Materials and Labor	Installed Cost
700	Wash Water *	10.04	8.88	18.92
	Waste-Water System **	12.70	7.45	20.15
800	Storage and Gasoline Blending	8.42	4.95	13.37
900	Steam and Power Generation *	120.52	47.82	168.34
	Energy Distribution *	15.06	5.33	20.39
	F.G. Desulfurization #	46.68	64.12	110.80
	Cooling Tower *	13.68	7.70	21.38
	Feed Water System *	17.58	7.02	24.60
	Ancilliary Units **	29.50	16.94	46.44
	Offsites Facilities **	135.96	82.59	218.55
	SUB-TOTAL	1,236.64	719.81	1,956.45
	Adjustment Factor for U.S. and Canadian Content (0.25 x 1.1592 +0.75 =1.0398)	49.22	-	49.22
	TOTAL	1,285.86	719.81	2,005.67

* Calculated from Krupp-Koppers' data:
** Calculated from ORNL report

# Calculated from NCB data.
All other data from Mobil's report.

Sources:

H.Staege, Krupp-Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982.
M. Schreiner, Mobil Research and Development, Corporation, DOE Report No. FE-2447-13, August 1978.
R. Wham et al, Oak Ridge National Laboratory, Report No. ORNL-5664, February 1981.
W.P. Syrett, NCB Report to SNC Inc, NCB Ref. No. DR-496/2, 7/3, January 23, 1982.

Table 6-7 Breakdown of Total Capital Investment of a SASOL-Type Plant via Fischer-Tropsch Synthesis (Synthol Reactor), in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of "as-received" Willow Bunch lignite]

Installed-Plant Cost (including process contingencies)	2,005.67
Project and Estimation Contingencies	200.57
Total Plant Installation Cost (TPIC)	2,206.24
Initial Catalysts, Chemicals and Operating Supplies	1.55
Start-up and Other Costs	66.19
Working Capital	66.19
Land	0.10
Interest During Construction	205.35
Total Capital Investment, "C"	2,545.62

Sources:

H. Staege, Krupp-Koppers GmbH report to SNC Inc., VK-1/Mn-Ko-1813, June 1982.

M. Schreiner, Mobil Research and Development Corporation, DOE Report FE-2447-13, August 1978.

B.C. McBeath, ESCOE, DOE Report FE-2468-44, April 1978.

6.2.3 **Coal-Based Plant for Production of Gasoline and Other Hydrocarbons via Fischer-Tropsch Synthesis** (cont'd)

Table 6-7. The total investment includes installed plant cost for all equipment items, bulk material, labor, engineering and support, licenses and prepaid royalties, process contingencies, project and estimation contingencies, land, working capital, interest during construction, etc. These costs have been calculated following the criteria and procedure described earlier.

The total capital investment does not include items such as coal-mine development costs; transportation facilities from mines to the plant; railways, roads and other transportation facilities outside the plant; water supply and pipeline costs outside the plant; housing, community facilities and associated infrastructure.

Table 6-8 Annual Operating, Maintenance and Production Costs of the Indirect Coal Liquefaction Plants, in Millions of Canadian Dollars (mid-1982)

[Capacity: 25,000 t/d of "as-received" Willow Bunch lignite]
[310 operating days per year]

Coal Based Plant Employing Koppers-Totzek Gasifier Configuration for	Methanol Production	Gasoline via Mobil-MTG	Gasoline and Hydrocarbons via F-T Synthesis
Fixed Expenses			
Operating Labour*	6.57	6.80	9.22
Operating Supplies	1.97	2.04	2.77
Supervisory Labour*	1.80	1.86	2.46
Labour Burden	2.93	3.03	4.09
Maintenance	50.06	57.53	66.19
Property Taxes and Insurance	45.89	52.74	60.67
Plant Overhead	24.16	27.29	32.26
Variable Expenses			
Coal Feed	75.62	75.62	75.62
Catalysts and Chemicals	14.50	15.70	16.65
Make-up Water	3.74	3.77	3.69
Waste Disposal	2.12	2.12	2.12
Running Royalties	-	-	-
Total Annual Operating Costs	229.36	248.50	275.74
Less Byproduct Credits	2.62	2.42	2.42
Total Annual Net-Operating Costs	226.94	246.08	273.32
Capital Costs**	220.64	253.56	292.68
	447.58	499.58	566.00

* Does not include maintenance labour and supervision as these items are covered in maintenance cost. Similarly, plant management and administrative staff are included in plant overhead.

** Capital cost factor "k" is 0.11457 for methanol plant, 0.11456 for MTG case, and 0.11497 for plant based on Fischer-Tropsch synthesis

Sources:

H. Staege, Krupp-Koppers report to SNC Inc., VK-1/Mn-Ko-1813, June 1982.
B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979.
R. Salmon et.al., Oak Ridge National Laboratory, Report No. ORNL/TM-7581, April 1981.
M. Schreiner, Mobil Research and Development Corporation, DOE Report FE-2447-14, August 1978

6.3 OPERATING, MAINTENANCE AND PRODUCTION COSTS

Operating, maintenance and production costs for the three indirect coal liquefaction process alternatives are presented in Table 6-8. These costs are reported on an annual basis assuming a plant service factor of 85%, i.e., 310 operating days per year at 100% design capacity (25,000 t/d of coal).

Operating and supervisory labour costs are calculated using a typical manpower distribution and average salary scale given in Appendix E-6. The total manpower estimated by Krupp-Koppers for methanol plant is 292. For Mobil-MTG plant, an additional* 10 operators are required raising the total to 302. For plant based on Fischer-Tropsch synthesis, an additional* 108 operators are required raising the total to 410.

Operating supplies are estimated at 30% of operating labour cost. A labour burden totalling 35% of direct labour cost (operating and supervisory) is added to cover fringe benefits, sick leave, social security, unemployment insurance and shift differentials, etc.

Maintenance is estimated at 3% of total plant installation cost (TPIC). This includes maintenance materials, replaced equipment items, maintenance labour, supervision, labour burden and other regular maintenance contracts.

Property taxes and insurance are assessed at 2.75% of the total plant installed cost (TPIC) comprising 2.0% local taxes and 0.75% insurance.

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* M. Schreiner, Mobil Research and Development Corporation, DOE Report FE-2447-13, August 1978.

Plant overhead is assumed to be 40% of the operating and supervisory labour, maintenance and operating supplies. This cost includes plant management, engineering, fire protection, medical services, personnel, accounting, clerical and janitorial services.

Costs of coal, catalyst and chemicals, utilities and waste disposal are treated as variable expenses. Cost of coal (Willow Bunch lignite) supplied to the plant is $9.75 per tonne. This price was provided by Saskoil* and includes $0.28 per tonne transportation costs from the mine to the plant.

Raw water is the only utility purchased from an outside source and is available to the plant site at a cost of $1.10 per 1000 imperial gallons. This cost has been estimated by SNC-Keith-Gabriel, Saskatchewan and includes costs of canal and water pumping from Diefenbaker Lake.

Waste disposal cost (for ash) is assumed to be $2/tonne.

Running royalties are payment on an annual or semi-annual basis to the process developer's licensors for the use of a proprietary process. The nature of royalty payment is complex and requires more detailed information than that possible within the scope of this assessment. Therefore, the running royalties have been neglected.

Byproduct credits for sulfur at $65/t, NH_3 at $300/t and mixed phenols at $250/t have been taken into consideration. The credit for CO_2 byproduct has not been considered due to market uncertainty.

F. Mourits, Coal Liquefaction Study Management, Saskoil, letter to J.A. Mikhlin, SNC Inc., Dec 17, 1981.

Levelised capital costs and investment related charges are calculated per annum in "constant" dollars as shown in Table 6-8. In the first and second years of operation, the plant is operating at 50% and 90% of design capacity respectively. To account for deficits in these two years, bridge financing at 10% interest rate in "real" terms and amortized over the balance period (28 years) is assumed. From the third year of operation, the plant is expected to operate at 100% design capacity, i.e., processing 25,000 t/d of coal.

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6.4 PRODUCT COST

Product costs for each of the three indirect coal liquefaction process alternatives is presented on "Energy Value" and "Product Value" bases. Product cost on "Energy Value" basis is calculated considering heating value only, whereas the "Product Value" also takes into account the relative market value of individual products. The product costs calculated for both "Energy Value" and "Product Value" bases are for relative ranking and selection of the process and are not intended for comparison with existing market values.

The first approach, product cost on an "Energy Value" basis, is simply the unit energy cost ($\$/10^6$kJ) and is obtained by dividing the total annual costs by the total energy produced. The second approach was proposed and used by The Engineering Society Commission on Energy (ESCOE)*. The objective of the ESCOE method is to allow comparison to be made both on product quality as well as heating value. For example, gasoline has a higher market value than fuel oil and calculating the product cost solely on heating value would invert the price ranking of these fuels since heating oil has a higher heating value per litre than does gasoline.

In the ESCOE methodology, a value factor F_i is defined as the ratio of market value of an individual product M_i to the chosen reference product M_r, thus:

$$F_i = \frac{M_i}{M_r} = \frac{P_i}{P_r} \qquad (1)$$

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* K.A. Rogers and R.F. Hill, "Coal Conversion Comparison", DOE Report No. FE-2468-51, July 1979.

where, M_i = market value of individual product
M_r = market value of the reference product
P_i = cost of individual product, \$/10^6 kJ
and P_r = cost of reference product, \$/10^6 kJ

On an annual basis, a simplified overall cost equation may be represented as:

$$\sum P_i B_i = F + V + kC \qquad (2)$$

where, P_i = cost of individual product, \$/10^6kJ
B_i = annual production (energy output) of product "i" in 10^6 kJ
F = annual fixed cost
V = annual variable cost
k = levelized capital cost factor to allow for interest, taxes, return and recovery of capital investment

C = total capital investment.

Substituting the value of P_i from Equation (1) into Equation (2) gives:

$$P_r = \frac{F + V + kC}{\sum F_i B_i} \qquad (3)$$

Equation (3) is the basic cost equation relating the cost of a reference product P_r as a function of annual costs, annual production and relative market values of various products. Thus, when P_r is determined, the cost of individual products on a "Value Basis" is obtained from Equation (1).

6.4 PRODUCT COST (cont'd)

To calculate product cost P_r, a set of market values M_i is required for each individual product to calculate value factor F_i relative to premium gasoline M_r. The market values supplied by Petro-Canada are shown in Table 6-9. The market values are wholesale price and include Federal Excise Tax. Provincial sales tax is not included.

The product cost on an "Energy Basis", P_e is calculated by assigning the value of $F_i = 1.0$ for all the products in Equation (3) to obtain:

$$P_e = \frac{F + V + kC}{\sum B_i} \qquad (4)$$

where, $\sum B_i$ = total annual production in 10^6 kJ.

Table 6-10 presents the production rate and annual energy output for the three indirect coal liquefaction process alternatives. Using the operating, maintenance and production costs in Table 6-8, the product costs on "Energy Value" and "Product Value" bases are calculated and presented in Table 6-11.

Table 6-9 Value Factor of Products from Indirect Coal Liquefaction Plants

(Reference to Premium Gasoline in Saskatchewan)

Products	Market Value *	Converted Market Value**, $/10^6 kJ	Value Factor F_i
Fuel Gas	-	-	0.35 (assumed)
Propane	$14.80/bbl	4.82	0.40
Propane/Butane (LPG)	$19.00/bbl	5.01	0.42
Butane	$23.24/bbl	5.98	0.58
No.2 Fuel Oil	$29.30/bbl	7.34	0.61
Diesel Fuel	$0.311/L	9.91	0.82
Unleaded Gasoline	$0.344/L	11.54	0.96
Mixed Alcohols	-	-	0.96 (assumed)
Methanol	$40.87/bbl	13.81	1.28
Premium Gasoline (Reference Product)	$0.359/L	12.04	1.00

* Market Values include Federal Excise Tax but Provincial Sales Tax is excluded. Price for diesel, unleaded gasoline, and premium gasoline are wholesale price for Saskatoon, Saskatchewan.

** Market value supplied by PETRO-CANADA have been converted to a consisted unit basis ($/10^6 kJ) using conversion data given by ESCOE, DOE Report FE-2468-51, July 1979.

Source:

M. Taherzadeh, Coal and Alternate Fuels, PETRO-CANADA, Telex to J.A. Mikhlin, SNC Inc., 7 April 1982

K.A. Rogers and R.F. Hill, ESCOE, DOE Report FE-2468-51, July 1979.

Table 6-10 Production Rate and Annual Energy Output from the Indirect Coal Liquefaction Plants

[Capacity: 25,000 t/d of Willow Bunch lignite]

[310 operating days per year]

Coal Based Plant	Product	Flow rates bbl/d	Flow rates t/d	Heating Value kJ/kg	Energy Output, 10^{12} kJ/year $\sum B_i$	Energy Output, 10^{12} kJ/year $\sum F_i B_i$
for Methanol Production	Pure Methanol	43,761	5,714	22,670	40.19	51.44
via Mobil MTG Process	Fuel Gas	-	64	37,643	7.46	2.61
	Propane	1,456	117	50,009	1.81	0.72
	Butane	2,062	186	48,940	2.83	1.64
	Gasoline	19,285	2,245	46,276	32.20	30.91
					44.30	35.88
via Fischer-Tropsch Synthesis (Synthol Reactor)	Fuel Gas	-	784	62,460	15.16	5.31
	Propane	855	69	49,923	1.05	0.40
	Butane	117	11	50,395	0.16	0.09
	Heavy Fuel Oil	483	63	46,206	0.90	0.55
	Diesel	1,785	214	47,069	3.13	2.57
	Gasoline	9,340	1,055	46,915	15.35	14.74
	Alcohols	1,047	133	31,187	1.30	1.24
					37.05	24.91

Table 6-11 Product Costs for Indirect Coal Liquefaction Plants

[mid-1982 Canadian dollars]

		Methanol Plant	Mobil MTG	Sasol-Type F-T plant
Average Product Cost, $\$/10^6$kJ				
Energy Value Basis		11.14	11.28	15.27
Product Value Basis*		8.70	13.92	22.72
Individual Product Cost (Product Value Basis), $/bbl				
Fuel Gas		-	14.75	40.08
Propane		-	22.37	36.61
Propane/Butane (LPG)		-	-	-
Butane		-	35.64	62.45
Heavy Fuel Oil		-	-	83.54
Jet Fuel		-	-	-
Diesel		-	-	105.11
Gasoline		-	69.40	115.57
Mixed Alcohols		-	-	86.40
Methanol		32.97	-	-
Individual Product Cost (Product Value Basis)				
Fuel Gas	$\$/10^6$kJ	-	4.87	7.75
Propane	$/bbl	-	22.37	36.61
Propane/Butane (LPG)	$/bbl	-	-	-
Butane	$/bbl	-	35.64	62.45
Heavy Fuel Oil	¢/L	-	-	83.54
Jet Fuel	¢/L	-	-	-
Diesel	¢/L	-	-	66.10
Gasoline	¢/L	-	45.30	72.70
Mixed Alcohols	$/bbl	-	-	86.4
Methanol	$/bbl	32.97	-	-

* With reference to premium gasoline in Saskatchewan

REFERENCES AND SOURCES

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REFERENCES AND SOURCES

Overview of Indirect Coal Liquefaction Processes

F. Boegner, Davy International AG, T.J. Kendron, T.K. Subramanian, Davy McKee Corp. "Winkler Fluid Bed Coal Gasification Experience With Low Grade Coals", proceedings of Seventh Annual International Conference on Coal Gasification, Liquefaction and conversion to Electricity, held at University of Pittsburgh, Pittsburgh, PA, August 5-7, 1980, pp. 217.

W. Koog, Texaco Development Corporation, "The Texaco Coal Gasification Process- Status and Outlook", The Eight Annual WATTec Energy Conference and Exhibition, Knoxville, TN, February 18-20, 1981.

W. Adlhock and K.A. Theis, Rheinische Brannkohlen-Werke ag, "The Rheinbraun High-Temperature Winkler (HWS) Process/ EPRI Conference Proceedings. Synthetic Fuels-Status and Directions, Vol. 2: Section 18- Appendices EPRI WS-79-238, May 1981, San Francisco, CA pp 29.1

B. Cornils and R. Specks, Ruhrchemie AG, "Experiences with the Texaco Process of Coal-Dust Pressure Gasification using the Ruhrchemie/Ruhrkohle Technical Version". EPRI Conference Proceedings: Synthetic Fuels Status and Directions, Vol. 2: Section 18- Appendices, EPRI WS-79-238, May 1981, San Francisco, CA pp.19-1.

J.D. Holmgren, C.E. Seglem, L.A. Salvador, and M.W. Dyos, Westinghouse Electric Corporation, "The Westinghouse Gasification Program". EPRI Conference Proceedings: Synthetic Fuels - Status and Directions, Vol. 2: Section 18- Appendices EPRI WS-79-238, May 1981, SAN Francisco, CA pp. 28-1.

C. Lohmann and J. Langhoff, Ruhrkohle AG, "Ruhr-100 Development Work on the Pressurized Lurgi Gasifier in Dorsten" EPRI Conference proceedings: Synthetic Fuels-Status and Directions Vol. 2: Section 18- Appendices EPRI WS-79-238, May 1981, San Francisco, CA pp. 23-1.

P.E. Rudolph, C. Hafke, and P.K. Herberg, Lurgi Kohle Und Mineralöltechnik GmbH "Lurgi Coal Gasification", EPRI Conference proceedings: Synthetic Fuel- Status and Directions, Vol. 2: Section 18- Appendices EPRI WS-79-238, May 1981, San Francisco, CA, pp. 22-1.

R.B. Sharman and J.E. Scott, British Gas Corp. "The British Gas/Lurgi Slagging Gasifier: Relevance to Power Generation". EPRI Conference Proceedings: Synthetic Fuel-Status and Directions Vol. 2, Section 18- Appendices, EPRI WS-79-238, May 1981, San-Francisco, CA, pp. 24-1.

F.B. Walter, H.C. Kaufman, Texaco Inc. and T.L. Reed, Southern California Edison Co. "The Cool Water Coal Gasification Combined Cycle Technology" EPRI Conference Proceedings: Synthetic Fuels-Status and Directions. Vol. 2: Section 18- Appendices, EPRI WS-79-238, May 1981, San-Francisco, CA, pp. 20-1.

REFERENCES AND SOURCES (cont'd)

Overview of Indirect Coal Liquefaction Processes (cont'd)

J.A. Mikhlin, SNC Inc., Visit to Lurgi Mineralöltechnick GmbH, Lurgi Research and Development Centre, October 29, November 5, 1981.

J.A. Mikhlin, SNC Inc., Visits to Rheinbraun on HTW Coal Gasification, cologne, Ruhrchemie/Ruhrkohle/Texaco Coal Gasification, Oberhausen, British Gas Corporation, on British Gas/Lurgi Gasifier UK, October 29-November 5, 1981.

J.A. Mikhlin, SNC Inc., Visit to Westinghouse Research and Development Centre and Waltz Mills Pilot Plant on Westinghouse Coal Gasification Process, January 1982.

H. Staege, Gesellschaft Für Kohle-Technologie mbH. "Entrained Bed Gasification" Hydrocarbon processing, March 1982, pp. 92-96.

J.A. Mikhlin, SNC Inc., Visit to Krupp-Koppers, Krupp Research and Development Centre, Essen, on Krupp-Koppers Coal Gasification process, April 1982.

W.B. Crouch, "The Texaco Coal Gasification Process - Synthesis Gas for Chemical Feedstocks" International Coal Conversion Conference, August 1982, Pretoria, South Africa.

M. Ilgner, K.H. Lim and E. Nitschke "Considerations for the Design of a Commercial Coal Gasification Plant According to the Texaco Process" International Coal Conversion Conference August 1982, Pretoria, South Africa.

P.E. Rudolph and C.P. Hafke "Lurgi Coal Gasification- Achievement and Developments" International Coal Conversion Conference, August 1982, Pretoria, South Africa.

L.A. Salvador and P.J. Meyer "The Joint SALSOL/Westinghouse Commercial Scale Fluidized Bed with a Draught Tube" International Coal Conversion Conference, August 1982, Pretoria, South Africa.

H. Teggers, W. Adlhoch, V. Femmer, and K.A. Theis "Actual Status of the Rheinbraun High-Temperature Winkler (HTW) Gasification Process" International Coal Conversion Conference, August 1982, Pretoria, South Africa.

Techno-Economic Assessment of Indirect Coal Liquefaction Processes

W.C. Hueper, "Experimental Carcinogenic Studies on Hydrogenated Coal Oils. II, Fischer-Tropsch Oils", Industrial Medecine and Surgery, Vol. 25, pp. 459-62, 1956.

J.C. Hoogendoorn, "Experience with Fisher-Tropsch Synthesis at SASOL", Institute of Gas Technology Conference, Chicago, September 1973.

REFERENCES AND SOURCES (cont'd)

Techno-Economic Assessment of Indirect Coal Liquefaction Processes (cont'd)

J.H. Govaarts and C.W. Schutte, "Commercial Evaluation of Fischer-Tropsch Process," South African Chemical Processing, November 1973.

M. Gibbs, "Sasol Technicians Arouse World Interest", South African Chemical Processing, November 1973.

"Development Document for Effluent Guidelines and New Source Performance Standards for the Petroleum Refining Point Source Category", U.S. Environmental Protection Agency, EPA Report No. 440/1-74-014A, Washington, D.C., April 1974.

Environment Saskatchewan, Air Pollution Control Regulations, 211/75, 1975.

Environment Saskatchewan, Water Pollution Control Branch, "Water Quality Objectives, January 1975, February 1977 Edition.

S.L. Meisel, "Gasoline From Methanol in One Step", CHEMTECH, February 1976

M.E. Dry, "Advances in Fischer-Tropsch Chemistry", Ind. Eng. Chem., Prod. Res. Dev., Vol. 15, No.4, 1976.

N. Daviduk, J. Maziuk, J.J. Wise, "Design of a 100 BPD Pilot Plant to Convert Methanol to Gasoline using Mobil Process", 11th Intersociety Energy Conversion Engineering Conference, 1976.

J.B. O'Hara et al., The Ralph M. Parsons Company, "Fischer-Tropsch Complex: Conceptual Design/Economic Analysis. Oil and SNG Production", R & D Report No. 114 - Interim Report No. 3, Energy Research and Development Administration, Washington, D.C., January 1977.

J.C. Hoogendoorn, "Gas From Coal For Synthesis of Hydrocarbons", Panel Discussion, Ninth Synthetic Pipeline Gas Symposium, Chicago, Oct. 3-Nov. 2, 1977.

Saskatchewan Power Corporation, "Research and Development Centre, Coronach Coal Gasification Study", Volume II, Job. No. C8093, October 30, 1977.

B.I. Loran and J.B. O'Hara, The Ralph M. Parsons Company, "Environmental Factors for Fisher-Tropsch Coal Conversion Technology", presented in Environmental Aspects of Fuel Conversion Technology III, pp. 855-863 September 1977, Hollywood, Florida. Symposium Proceedings compiled by F.A. Ayer and M.F. Massoglia of Research Triangle Institute for the U.S. Environmental Protection Agency, EPA Report 600/7-78-063, April 1978.

REFERENCES AND SOURCES (cont'd)

Techno-Economic Assessment of Indirect Coal Liquefaction Processes (cont'd)

M. Schreiner, Mobil Research and Development Corporation, "Research Guidance Studies to Assess Gasoline from Coal by Methanol-to-Gasoline and SASOL-type Fisher-Tropsch Technologies", report prepared for the U.S Department of Energy under Contract No. EF-76-C-01-2447, Report No. FE-2447-14, August 1978.

B.C. McBeath, The Engineering Societies Commission on Energy Inc., "Guidelines for Economic Evaluatuion of Coal Conversion Processes", prepared for the U.S. Department of Energy under Contract No. EF-77-C-01-2468, Report No. FE-2468-44, April 1979.

K.A. Rogers and R.F. Hill, The Engineering Societies Commission on Energy Inc., "Coal Conversion Comparison, prepared for the U.S. Department of Energy under Contract No. EF-C-01-2468, Report No. FE-2468-51, July 1979.

J.C.Hoogendoorn, "Motor Fuels and Chemicals From Coal via The SASOL Synthol Route", Paper presented at Royal Society, London, May 21, 1980.

D.Gray et al., "The Impact of Developing Technology on Indirect Liquefaction", Report Prepared for the U.S. Department of Energy by Mitre Corp., Va., DOE/ET/10280-T_1T_2, November 1980.

C.A. Zee et al., TRW Inc., California, "Environmental Assessment: Source test and Evaluation Report on Koppers-Totzek Process", EPA Report No. 600/7-81-009, Janurary 1981.

R.M. Wham et al., Oak Ridge National Laboratory, "Liquefaction Technology Assessment - Phase 1: Indirect Liquefaction of Coal to Methanol and Gasoline Using Available Technology", Oak Ridge National Laboratory Report ORNL-5664, February 1981.

K.A. Rogers, U.S. Synthetic Fuels Corp. "Synfuel Technologie and Costs", paper presented at the AIME Meeting, Chicago, Illinois, February 24, 1981, published in Material and Society, Vol. 5, No. 4, pp. 353-364, 1981.

R. Salmon et al., Oak Ridge National Laboratory, "Technical and Economic Evaluation of Coal Liquefaction Technologies. Phase 0: Preliminary Screening Evaluation", Report No. ORNL/TM-7581, April 1981.

J.C. Hoogerdoorn, "Production of Automotive Fuels From Coal in South Africa", American Petroleum Institute, Mid-Year Meeting, Refining Department, Chicago, May 11-14, 1981.

REFERENCES AND SOURCES (cont'd)

Techno-Economic Assessment of Indirect Coal Liquefaction Processes (cont'd)

Construction Price Statistics, "Chemical and Mineral Process Plant Indexes", Table 19; and "Union Wage Rate Index", Table 3; November 1981, Catalogue No. 62-007, published by Statistics Canada. Ottawa.

Engineering News Record, January 7, 1982; January 16, 1982.

W.P. Syrett, National Coal Board, "The Processing of Saskatchewan Lignite Report to SNC Inc., by the NCB's Liquid Solvent Extraction Process: Willow Bunch Lignite", NCB Reference No. 496/2,7/3, January 23, 1982. (The report is attached in confidential Volume V of the present study, "Process Developers Data and Information").

F. Mourits, Saskoil, letter to J. A. Mikhlin, December 15, 1982.

M. Taherzadeh Coal and Alternate Fuels, PETRO-CANADA, telex to J.A. Mikhlin, SNC Inc., 7 April 1982.

Construction Collective Agreement Summaries, "Labour Rate in Saskatchewan", published by the Canadian Construction Association, Ottawa, April 1982.

H. Staege, Krupp Koppers GmbH, "Preliminary Study for the Production of Pure Methanol or Synthesis Gas from Coal on the Basis of the Kopper-Totzek Coal Gasification Process", Report to SNC Inc., Krupp-Koppers Ref. No.VK-1/Mn-Ko-1813, June 1982. (The report is attached in the confidential Volume V of the present study, "Process Developers Data and Information").

APPENDIX A

ULTIMATE, PROXIMATE AND PETROGRAPHIC ANALYSIS OF SASKATCHEWAN LIGNITES

Table A-1 Typical Ultimate, Proximate, and Petrographic Analysis of Saskatchewan Lignites

	Estevan	Willow Bunch
Moisture (as received), wt%	35.0	35.0
Gross Calorific Value (GCV) kJ/kg MF coal	24,270	20,270
Proximate Analysis, wt% MF coal		
Volatile Matter	42.0	39.6
Fixed Carbon	42.0	38.5
Ash	15.4	21.0
Sulphur	0.6	0.9
Ultimate Analysis, wt% MF coal		
Carbon	61.2	52.1
Hydrogen	4.5	4.0
Nitrogen	1.0	0.5
Sulfur	0.6	0.9
Oxygen	17.3	21.5
Ash	15.4	21.0
Petrographic Analysis, vol%		
Vitrinite*	81.2	77.6
Exinite	7.9	1.7
Inertinite	9.9	18.7
Mineral Matter	0.9	2.0
Ash Analysis, wt%		
Silica, SiO_2	47.1	42.1
Alumina, Al_2O3	17.8	15.6
Titania, TiO_2	**	0.9
Ferric Oxide, Fe_2O_3	2.5	6.7
Lime, CaO	11.7	14.3
Magnesia, MgO	2.9	5.0
Potassium Oxide, K_2O	0.5	1.3
Sodium Oxide, Na_2O	5.0	1.8
Phosphorous Pentoxide, P_2O_5	1.0	
Sulfur Trioxide, SO_3	1.0	12.3
Undetermined	10.5	

* Equivalent to huminite in low-rank coals

** Undetermined.

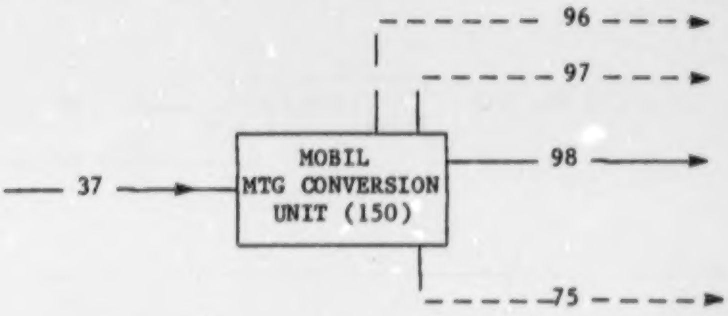
Source: Information package received from Saskoil at the Project Kick-off Meeting, Sept. 10-11, 1981

Letter from F.M. Mourits, Coal Liquefaction Study Management, Saskoil, dated December 15, 1981 to J.A. Mikhlin, SNC Inc.

APPENDIX B

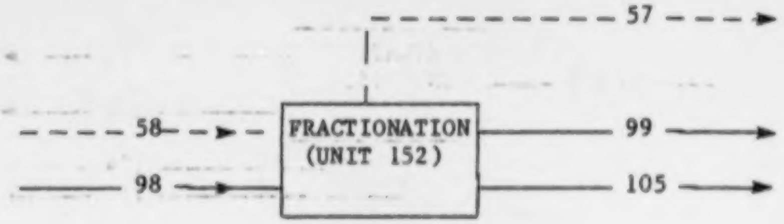
MATERIAL BALANCE OF METHANOL-TO-GASOLINE PROCESS

Table B-1 Material Balance - Methanol Conversion (Unit 1500)



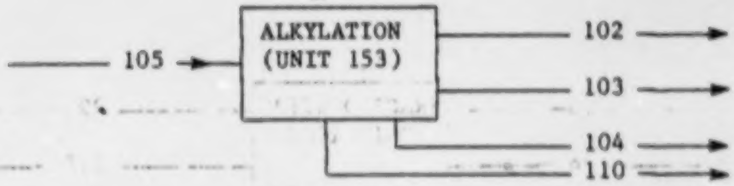
STREAM NUMBER	37	75	96	97	98
DESCRIPTION	CRUDE MeOH	SOUR WATER TO SECTION 700	PURGE GAS TO ATMOSPHERE	FUEL GAS	WILD GASOLINE TO FRACTIONATION UNIT
Components:					
kg/d					
CO_2	14,352	2,820		1,416	14,493
CO	54	31		272	642
H_2	6			54	98
CH_4	1,307			1,517	20,221
C_2H_4				53	893
C_2H_6	610				10,856
N_2	8			6	1
H_2O	253,938	3,606,359		130	2,352
MeOH	5,999,031				
Acetone		14,047			
Fatty Acid		14,058			
C_3H_6				98	4,955
C_3H_8				2,090	119,739
iC_4H_{10}				2,068	232,704
iC_4H_8				222	28,171
nC_4H_{10}				490	71,956
iC_5H_{12}				1,081	317,151
iC_5H_{10}				185	57,005
nC_5H_{12}				103	36,142
C_6		4,312		909	1,694,548
Coke			3,055		
TOTAL(kg/d)	6,269,304	3,641,627	3,055	12,694	2,611,927

Table B-2 Material Balance - Wild Gasoline Fractionation (Unit 152)



STREAM NUMBER	58	98	57	99	105
DESCRIPTION	LIGHT GAS CONDENSATE FROM SNG TRAIN	WILD GASOLINE FROM UNIT 150	FUEL GAS	STABILIZED GASOLINE TO BLENDING UNIT	ALKYLATION FEED
Components:					
kg/d					
CO_2		14,493	14,450		
CO		642	645		
H_2		98	98		
CH_4		20,221	20,222		
C_2H_4		893	893		
C_2H_6		10,856	10,695		294
N_2		1	5		
H_2O		2,352	2,352		
MeOH					
Acetone					
Fatty Acid					
C_3H_6		4,955	291		4,661
C_3H_8		119,739	6,000	4	113,728
iC_4H_{10}		232,704	4,737	4,997	222,980
iC_4H_8		28,171	487	1,378	26,309
nC_4H_{10}		71,956	1,045	7,775	63,136
iC_5H_{12}		317,151	1,608	310,105	5,434
iC_5H_{10}		57,005	214	56,350	436
nC_5H_{12}		36,142	103	36,018	20
C_6		1,694,548	131	1,694,327	
TOTAL(kg/d)	0	2,611,927	63,976	2,110,953	436,998

Table B-3 Material Balance - Alkylation (Unit 153)



STREAM NUMBER	105	102	103	104	110
DESCRIPTION	ALKYLATION FEED	PROPANE LPG PRODUCT	BUTANE LPG PRODUCT	ALKYLATE TO GASOLINE BLENDING UNIT	MIXED BUTANES TO GASOLINE BLENDING UNIT
Components:					
kg/d					
CO_2					
CO					
H_2					
CH_4					
C_2H_4					
C_2H_6	294	294			
N_2					
H_2O					
MeOH					
Acetone					
Fatty Acid					
C_3H_6	4,661				
C_3H_8	113,728	114,447			
iC_4H_{10}	222,980	2,505	141,140	1,390	43,670
iC_4H_8	26,309	9			
nC_4H_{10}	63,136		43,825	5,811	13,558
iC_5H_{12}	5,434		1,355	6,345	416
iC_5H_{10}	436		71	345	24
nC_5H_{12}	20			24	
C_6				61,769	
TOTAL(kg/d)	436,998	117,255	186,391	75,684	57,668

snc

APPENDIX C

FISCHER-TROPSCH CATALYST PREPARATION

Appendix C: Fischer-Tropsch Catalyst Preparation

A promoted reduced iron catalyst for the fluidized-bed Fischer-Tropsch Synthol reactor is prepared from magnetite iron ore (Fe_3O_4). The details of catalyst preparation technique such as types and amounts of promoters used, concentrate specifications and specific operating conditions are confidential.

The description given below is based on review of literature sources: (i) "The Manufacture of Iron Ammonia Catalyst", CEP, 43, pp 291, June 1947, and (ii) "The SASOL Catalyst Plant", Brit. Chem. Engr., May, June, July, Aug. 1957".

Catalyst preparation consists of 3 steps: iron ore purification, catalyst fusion and size reduction, and catalyst reduction. Equipment used for iron ore purification includes:

- ore unloading and storage
- rotary drier for moisture removal before grinding
- rod mill for size reduction
- air separation table for tailings removal
- concentrate storage

Equipment used for catalyst fusion and size reduction step includes:

- electric resistance furnace for melting and intimate mixing of concentrate and promotors
- gyratory crusher and ball mill for size reduction per fluidized-bed reactor requirements
- air separator for elimination of fines and oversize particles
- catalyst storage

Equipment required for catalyst reduction includes:

- fluidized-bed reactor for contact of hydrogen and unreduced F-T catalyst
- hydrogen heater
- heat exchangers for condensing water
- hydrogen recycle compressor
- recycle gas drier for moisture control
- reduced catalyst storage under H_2 or inert gas atmosphere

Reduction is carried out under elevated temperature and pressure.

In terms of yields, approximately 30 t/d of reduced catalyst can be produced by processing 91 t/d of iron ore.

APPENDIX D

REGULATORY ENVIRONMENT CONTROL ACTS
PROVINCE OF SASKATCHEWAN

Appendix D: Regulatory Environmental Control Act - Province of Saskatchewan

The proposed lignite liquefaction plant in the province of Saskatchewan will have to comply with the applicable regulations under the following acts:

1. THE AIR POLLUTION CONTROL ACT

This act requires that permits are secured for operating fuel-burning equipment which certainly would pertain to any of the lignite liquefaction route (direct or indirect) under consideration. The Air Pollution Control Regulations contains ambient air quality standards which are summarized in Table D-1 for the air pollutants of suspended particulates, sulfur dioxide and nitrogen dioxide.

2. THE WATER RESOURCES MANAGEMENT ACT

The act requires that construction and operational permits are obtained in respect of proposed water and sewage works or polluted effluent discharges. The Water Pollution Control and Waterworks Regulations specify the permit application requirements. Guidelines regarding the design of water and sewage works are given in separate documents, which are exclusively pertaining to municipal works. Regarding liquid effluents from the lignite liquefaction plant under consideration, it is to be expected that project specific water pollution requirements would be defined by Saskatchewan Environment during the course of a permit application. Some general policy statements and effluent discharge objectives are given in the Water Quality Objectives issued by Saskatchewan Environment. This document includes several numerical objectives which are summarized in Tables D-2 through D-5.

3. THE WATER RIGHTS ACT

The act requires that approvals are secured in terms of surface water use, obstruction, diversions or storage. These provisions would possibly apply to the cooling system water supply chosen for any of the direct or indirect lignite liquefaction plant.

4. THE GROUND WATER CONSERVATION ACT

The act requires that permits and licenses be secured prior to the exploration for and use of ground waters respectively. The Ground Water Regulations specify the requirements for obtaining a permit or license. The legislation also contains provisions for prevention of ground water pollution. The possible use of ground water in the direct or indirect lignite liquefaction plants as well as the ground water dewatering systems for the proposed mine would require several approvals under this act.

5. THE PUBLIC HEALTH ACT

The act mostly contains provisions regarding hygienic matters. The Waste Management Regulations of 1972 may however apply to the disposal of ash and slag rejects from the lignite liquefaction plant. The regulations stipulate that a permit be secured in respect of the location, operation and maintenance of solid waste disposal systems. The regulations also contain provisions for cooperation between the Public Health and Environment Departments in terms of setting directives for the disposal of toxic wastes or substances.

Table D-1 Ambient Air Quality Standards - Province of Saskatchewan

Air Contaminant	Unit	Average Concentration for Applicable Time Period				
		1 hour	8 hours	24 hours	30 days	1 year
Suspended Particulates	ug/m^3	-	-	120	-	70 *
Settable Particulates	mg/cm^2	-	-	-	2	-
Soil Index	COH units	-	-	1.5	-	-
Potash	mgK+/cm^2	-	-	-	0.15 or	-
	mgCl$^-$/cm^2	-	-	-	0.15	-
Sulphur Dioxide ***	ug/m^3	450	-	150	-	30 *
	ppm (volumetric)	0.17	-	0.06	-	0.01
Sulphation	mg/100 cm^2	15	-	-	30	-
Carbon Monoxide	mg/m^3	15	6	-	-	-
	ppm (volumetric)	13	5	-	-	-
Oxidents (Ozone)	ug/m^3	100	-	30	-	-
	ppm (volumetric)	0.05	-	0.015	-	-
Nitrogen Dioxide	ug/m^3	400	-	200	-	100 **
	ppm (volumetric)	0.2	-	0.1	-	0.05

* Geometric means.
** Arithmetic means.
*** Montana ambient standards for SO_2: 0.25 ppm (1 hr.), 0.10 ppm (24 hr.), 0.02 ppm (annual).

NOTES: No hydrogen sulfide gas can be released to the atmosphere from any industrial source, fuel-burning equipment or incinerator. Opacity shall be less than 40% or not more shady than No. 2 on the Ringelmann Chart; discharges during start-up or shut-down, in excess of 40% opacity are permitted for a duration of less than 5 minutes.

SOURCE: Environment Saskatchewan, Air Pollution Control Regulations, 211/75, 1975.

Table D-2 Saskatchewan Surface Water Quality Objectives*

Parameter	Objectives
Bacteriology (Coliform Group)	- In water to be withdrawn for treatment and distribution as a potable supply or used for outdoor recreation other than direct contact, at least 90 percent of the samples (not less than five samples in any consecutive 30-day period) should have a total coliform density of less than 5,000 per 100 mL and a fecal coliform density of less than 1,000 per 100 mL.
	- In waters used for direct contact recreation or vegetable crop irrigation the geometric mean of not less than five samples taken over not more than a 30-day period should not exceed 1,000 per 100 mL total coliforms, nor exceed these numbers in more than 20 percent of the samples examined during any month, nor exceed 2,400 per 100 mL total coliforms on any day.
Physical Quality	
Dissolved Oxygen	- A minimum of 5 mg/L at any time.
Biochemical Oxygen Demand	- Dependent on the assimilative capacity of the receiving water, the BOD must not exceed a limit which would create a dissolved oxygen content of less than 5 mg/L.
Suspended Solids	- Not to be increased by more than 10 mg/L over background value.
pH	- To be in the range of 6.5 to 8.5 pH units but not altered by more than 0.5 pH units from background value.
Temperature	- Not to be exceeded by more than 3°C above ambient water temperature.
Odour	- The cold (20°C) threshold odour number not to exceed eight.
Colour	- Not to be increased more than 30 colour units above natural value.
Turbidity	- Not exceed more than 25 Jackson units over natural turbidity.

* Apply to surface waters, except in areas of close proximity to outfalls.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table D-2 Saskatchewan Surface Water Quality Objectives (cont'd)

Parameter	Objectives (Maximum Concentrations)
	mg/L
Organic Chemical	
Carbon Chloroform Extract (CCE) (includes Carbon Alcohol Extract)	0.2
Methyl Mercaptan	0.05
Methylene Blue Active Substances	0.5
Oil and Grease	Substantially absent, no irridescent sheen
Phenolics	0.005
Resin Acids	0.1
Inorganic Chemical	
Boron	0.5
Copper	0.02
Fluoride	1.5
Iron	0.3
Manganese	0.05
Nitrogen (Total Inorganic and Organic)*	1.0
Phosphorous as PO_4 (Total Inorganic and Organic)*	0.15
Sodium (as percent of cations)	30 - 75
Sulphide	0.05
Zinc	0.05
Toxic Chemicals	
Arsenic	0.01
Barium	1.0
Cadmium	0.01
Chromium	0.05
Cyanide	0.01
Lead	0.05
Mercury	0.0001
Selenium	0.01
Silver	0.05

Notes: The predominant cations of sodium, calcium and magnesium and anions of sulphate, chloride and bicarbonate are too variable in the natural water quality state to attempt to define limits. Nevertheless, in order to prevent impairment of water quality, where effluents containing these ions are discharged to a water body the permissible concentration will be determined by the Department in accordance with existing quality and use. Unless specified, parameter concentrations refer to "total" measurements.

* These objectives are presently under study and may require adjusting according to naturally occurring concentrations or conditions.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table D-2 Saskatchewan Surface Water Quality Objectives (cont'd)

Parameter	Objectives
Pesticides	- To provide reasonably safe concentrations of these materials in receiving waters, an application shall not exceed 1/100 of the 48-hour Tl_m. Consideration must also be given to sublethal effects such as taste and odour generation.
Radioactivity	- Gross Beta not to exceed 1,000 pCi/L. - Radium 226 not to exceed 3 pCi/L. - Strontium 90 uot to exceed 10 pCi/L.
Unspecified Substances	- Substances not specified herein should not exceed values which are considered to be deleterious for the most critical use as established by the Department.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table D-3 Saskatchewan Municipal Drinking Water Quality Objectives

Parameter	Objectives
Bacteriology	
Total Coliform	- At least 90 percent of the samples in any consecutive 30 day period should be negative for total coliform organisms and no one sample should have a Most Probably Number index of greater than 10 per 100 mL or 30 per 100 mL by membrane filter. Properly operated municipal waterworks should be free of coliform bacteria.
Nuisance Biological Organism	- Biological organisms in concentrations which may produce objectionable colour, taste, odour and turbidity, or which may release toxid metabolites, or which may harbour pathogens are undesirable in drinking water and should be kept below such concentrations as to prevent any undesirable effects.
Radioactivity	
Gross Alpha Activity	- Not to exceed 10 pCi/L.
Strontium 90	- Not to exceed 10 pCi/L.
Radium 226	- Not to exceed 3 pCi/L.
Gross Beta Activity	- Not to exceed 1000 pCi/L (in absence of Alpha).
Other Specific Nuclides	- Not to exceed 0.1 of the maximum permissible concentration in water adopted by the International Commission on Radiological Protection.
Physical Quality	
	- Water should not contain impurities that would be offensive to the sense of sight, smell or taste.
Colour	- Maximum of 15 units.
Threshold Odor Number	- Maximum of 3 units.
Turbidity	- Maximum of 5 units.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table D-3 Saskatchewan Municipal Drinking Water Quality Objectives

Parameter	Objectives (Maximum Concentrations) mg/L
Chemical Quality	
Alkalinity ($CaCO_3$)	500*
Chloride	250
Copper	1.0
Fluoride	1.5
Iron	0.3
Hardness ($CaCO_3$)	800*
Magnesium	200*
(Magnesium & Sodium) and Sulfate	1,000*
Manganese	0.05
Methylene Blue Active Substances	0.5
Organics (carbon chloroform and carbon alcohol extractibles)	0.2
Phenolics	0.001*
Sodium	300*
Sulphate	500
Total Dissolved Solids	1,500*
Zinc	5.0
pH (permissable range)	7.0 - 9.5
Toxic Chemical	
Arsenic	0.01
Barium	1.0
Boron	5.0
Cadmium	0.01
Chromium	0.05
Cyanide	0.01
Lead	0.05
Nitrates (including nitrites) as NO_3	40*
Selenium	0.01
Silver	0.05
Biocides**	absent

* These values have been modified from those of the Canadian Drinking Water Standard and Objectives, 1968.

** In general, biocides should be virtually absent or not detectable in any drinking waters. The presence of any biocide would require a special investigation and study.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

Table D-4 Saskatchewan Guidelines for the Chemical Water Quality Respecting Private Water Supplies

Constituent	Range of Concentration Satisfactory Quality	 Poor Quality	 Not recommended for Consumption	 Unsuitable for Use
	mg/L	mg/L	mg/L	
Total Dissolved Solids	100 - 1,500	1,500 - 3,000	3,000 - 4,000	GT 4,000
Total Hardness	50 - 500	500 - 1,000	1,000 - 2,000	GT 2,000
Total Alkalinity ($CaCO_3$)	50 - 500	500 - 1,000	1,000 - 1,500	GT 1,500
Chloride	LT 250	250 - 500	500 - 1,000	GT 1,000
Sodium	LT 300	300 - 500	500 - 1,000	GT 1,000
Sulphate	LT 400	400 - 800	800 - 1,200	GT 1,200
Nitrate	LT 40	40 - 300	GT 300	- - - - -
Iron	LT 0.3	0.3 - 1.0	- - - - - - - -	- - - - -
Manganese	LT 0.05	0.05 - 0.5	- - - - - - - -	- - - - -
pH (units)	7.0 - 9.5	6.0 - 7.0 and 9.5 - 10.0	- - - - - - - -	LT 5.5 and GT 10.5

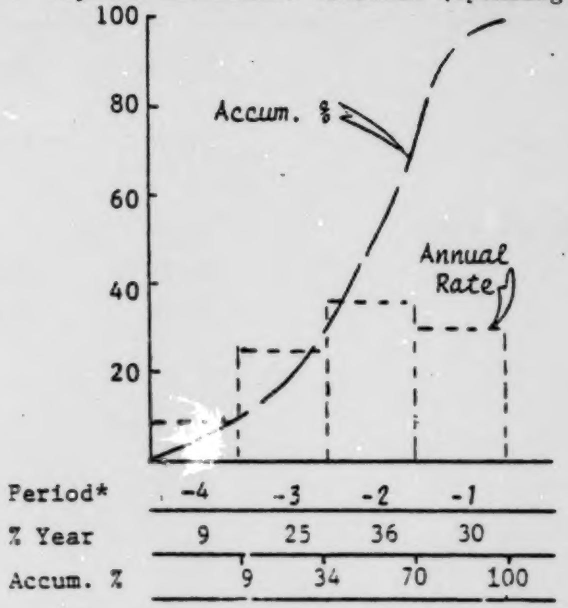
Abreviations, LT: Less than, GT: Greater than.

Source: Water Quality Objectives, Environment Saskatchewan, January, 1975.

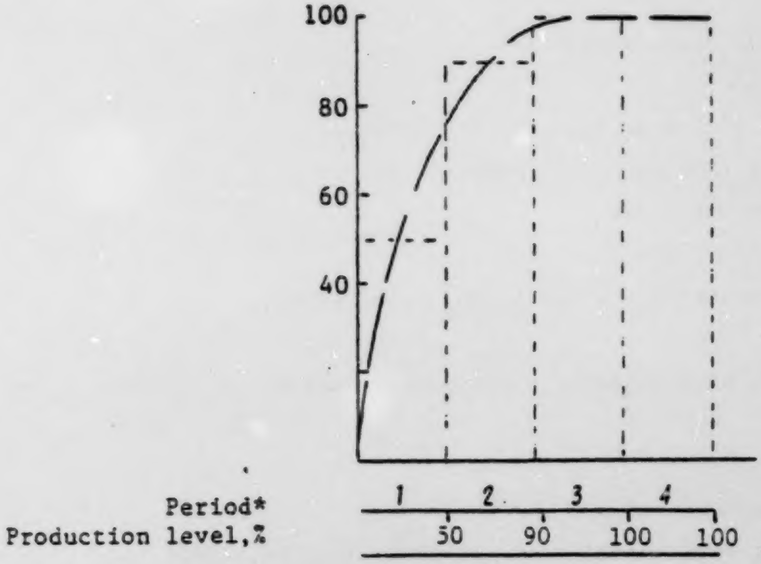
APPENDIX E

ECONOMIC AND FINANCIAL ASSUMPTIONS AND CRITERIA

Appendix E-1 Capital Investment Schedule (Spending Schedule)



Appendix E-2 Production Level (Percentage of Annual Production Rate)



With reference to plant start-up date.

Source: B.C. McBeath, ESCOE, DOE Report No. FE-2468-44, April 1979

Appendix E-3: Canada/U.S. Material Price Differences

	U.S.A.	Canada	Variance, %
Concrete Mix, U.S $/cwt	47.90	50.12	4.6
Standard Structures, U.S $/cwt			
Wide Flanges	26.20	30.95	18.13
I-Beams	27.70	33.95	22.60
Piping, U.S $/ft			
Iron	4.51	6.44	42.8
Copper	0.81	0.55	32.1
Cables/Wires			10.0

Assuming the material distribution of:

30% concrete
15% steel (50% wide flanges, 50% I-beams)
45% piping (80% iron, 20% copper)
10% cables and wires

the "Material Difference Factor" of 1.1592 is obtained.

Therefore, "Material plus Currency Differential Factor" is 1.1592 x 1.19
= 1.379

Source: Engineering News Record, Jan. 7, 1982; Jan. 14, 1982.

Appendix E-4: Conversion of Manhours to Saskatchewan (Productivity Adjustment Factor)

A review of labour productivity based on petrochemical projects indicates that a multiplier of 1.25 may be applied for conversion of U.S. Gulf Coast manhours to equivalent Canadian manhours.

For adjustment to Saskatchewan, the following two items must also be considered:

Unique Condition Factor

Projects of this size will generally experience during the construction period several occurances whereby labour will interpret certain management decisions as being an affront to their integrity and pride. It will be assumed that these types of non-strike situations will lead to a cumulative loss of one working day per month during the construction period.

Factor = one day/20 working days per month = 0.05

Climatic Factor

The Saskatchewan Region experiences considerably more sub-zero temperatures and the duration of the winter season is generally about two months longer than the Sarnia region. Therefore, an allowance of 20 days per winter will be made for adverse working conditions due to weather.

$$\text{Factor} = \frac{\text{20 days x 4 winters}}{\text{48 months x 20 days/winter}} = 0.08$$

Thus, the "Overall Productivity Adjustment Factor" is

= 1.25 x 1.05 x 1.08 = 1.417 = 1.42

Appendix E-5: Process Equipment and Plants Price Indexes

Year	Index	% Increase
1977	181.9	7.53
1978	195.6	10.70
1979	216.6	10.90
1980	240.2	11.25
1981	267.2	11.00
1982*	-	11.00

* Assumed that similar % increase will continue in 1982.

Source: Construction Price Statistics, "Chemical and Mineral Process Plant Price Indexes", Table 19, November 1981

Appemdox E-6: Typical Manpower Distribution and Average Salary Scale

	% of Total Manpower *	Average Salary $ Month **
Operating Labour (OL)	34	1575
Operation Supervisors (20% of OL)	7	2500
Maintenance Labour (ML)	42	1875
Maintenance Supervision (20% of ML)	8.5	2500
Plant Management and Administrative Support	8.5	2700

Sources:

* K.A. Rogers and R.F. Hill, ESCOE, DOE Report No. FE-2468-51 July 1979

** Values reported in SPC Study, (Job No. C8093) "Coronach Coal Gasification Study", Volume II, October 1977, have been adjusted to mid-1982 dollars using salary index in Appendix E-7.

Appendix E-7: Salary Increases for Operating Costs Calculation (Location: Saskatoon)

Year	Index	% Increase
1976	204.0	10.25
1977	224.9	5.47
1978	237.2	6.62
1979	252.9	9.25
1980	276.3	7.38
1981	296.7	9.38
1982*	-	11.38

* Increase for 1982 assumed as 1981 + 2%

Source: Construction Price Statistics, "Union Wage Rate Index", Table 3, November 1981

SASKOIL COAL LIQUEFACTION STUDY

Review and Evaluation of Indirect Liquefaction Processes with the Following Gasifiers:

Lurgi Dry Bottoms
BGC/L Slagging Bottoms
Koppers-Totzek
Winkler
Texaco

November 1981

Foster Wheeler Synfuels Corporation
110 SOUTH ORANGE AVENUE, LIVINGSTON, NEW JERSEY

INTRODUCTION

INTRODUCTION

The objective of the present prescreening study is to compare various indirect coal liquefaction processes. The ultimate goal as to select the most promising processes for the specific type of coal available and the geographical area in which the facility is to be located.

The selection will certainly require the consideration of many factors. Some of these factors are possibly incompatible, as for example, the maximization of thermal efficiency as opposed to the maximization of liquid products.

It is also evident that a certain amount of subjective judgment will be present in this preliminary screening. This is due mainly to the approximate nature of the data developed at this stage.

Another consideration that will obviously enter into play is the state-of-the-art at the specific time frame of reference.

This allowed the generation of the basic data from which meaningful comparisons between the various cases considered can be made. The development of a common baseline for all cases was a fundamental requirement for this study.

However, for comparison purposes, it is the relative values for the data obtained that are of interest. The accuracy of the absolute values presented are of an approximate nature.

The basic assumptions for the present prescreening study are:

1. Coal: Lignite from the Estevan zone with a proximate and ultimate analysis as shown in other parts of this report.

   Feed to the Plant: 25,000 STPSD, as received.

2. Coal will be available at the Facility Battery Limits, being shipped from the mine by rail, truck or conveyors.

3. The Facility is a "grass roots" one and only coal, water, and air are required as major raw materials.

   It is assumed that the water requirements will be satisfied from wells or otherwise.

4. Ash/slag will be disposed onsite or offsite. Sufficient acreage will be made available for 20 years storage, if onsite disposal.

5. Product quality will be as produced in within Battery Limits, with no upgrading or secondary refining.

6. Electrical Power in excess of the Facility captive requirement will be available for export.

7. Plant on stream factor: 90%

Due to the non-confidential nature of the present report, use of published references to complement the in-house information was made.

The following cases are considered in this report:

Case I: Lurgi Gasifier (dry bottoms) with methanol synthesis followed by the [*]MRDC's methanol to gasoline process.

Case IA: Maximization of liquid fuels by reforming the SNG produced in Case I.

Case II: Lurgi Gasifier (dry bottoms) with Fischer-Tropsch synthesis.

Case IIA: Maximization of liquid fuels by reforming the SNG produced in Case II.

Case III: Winkler Gasifiers with methanol synthesis followed by the MRDC's methanol to gasoline process.

Case IV: Winkler Gasifier with Fischer-Tropsch synthesis.

Case V: Koppers-Totzek Gasifier with methanol synthesis and MTG (methanol to gasoline by MRDC process).

Case VI: Texaco gasifier with methanol synthesis and MTG.

Case VII: British Gas Corp./Lurgi slagging bottom gasifier (BGC/L) with methanol synthesis and MTG.

Case VIIA: Maximization of liquid fuels by reforming the SNG produced in Case VII.

Case VIII: BGC/L gasifier with Fischer-Tropsch synthesis.

Case VIIIA: Maximization of liquid fuels by reforming the SNG produced in Case VIII.

[*] Mobil Research and Development Corporation

A Saskatchewan lignite from the Estevan zone with the following proximate and ultimate analysis was used in this report:

Proximate Analysis

	Weight %
Volatile Matter	27.30
Fixed Carbon	27.30
Moisture	35.00
Ash	10.01
Sulfur	0.39

Ultimate Analysis

	Weight %
H	2.60
C	39.78
N	0.65
O	11.57
S	0.39
Ash	10.01
H_2O	35.00

A lignite from the Willow Bunch Coal Field with the following proximate and ultimate analysis was also considered in this report:

Proximate Analysis

	Weight %
Volatile Matter	25.74
Fixed Carbon	25.03
Moisture	35.00
Ash	13.65
Sulfur	0.58

Ultimate Analysis

	Weight %
H	2.61
C	33.86
N	0.32
O	13.98
S	0.58
Ash	13.65
H_2O	35.00

This coal has a higher ash and oxygen and lower carbon content than the Estevan coal.

This will result in lower yields of products than the Estevan lignite.. No attempt was made to estimate the actual yields, thermal efficiencies and other parameters for the Willow Bunch coal.

In general terms, it may be said that the expected products yields from the Willow Bunch would be about 85% of the ones for the Estevan lignite.

SELECTION CRITERIA

1. Marketability of Products

The possibility of marketing the various products obtainable in the cases here considered will eliminate those cases which: by reason of quality specification not meeting the market requirements, the absence of market, or the price at which they can be sold, renders these cases not feasible.

2. Cost of Products (Comparative)

The cost of the products is made-up of three main components:

A. Capital Cost
B. Operating Cost
C. Capital Charges

It has been found that on the average approximately the cost of coal is about 25% of the total product cost, the operating cost about 15% and the capital charges about 60¢.

The capital charges will vary according to the type of financing considered, which could be divided into three main categories: 100% equity, 100% debt financing such as municipal bonds, and part equity and part debt, common in public utility financing.

The principal items of the capital charges are depreciation and interest on the debt.

For capital intensive plants, such as the ones considered here, the capital charges become a substantial portion of the total operating cost.

The coal cost will remain the same for all cases considered, the operating cost will vary somewhat from case to case, but it is felt that these variations are of a minor consequence.

The main variable cost item is the capital cost charges which are directly related to the Plant Cost, which amounts to about 60% of the total products cost.

As a consequence, the comparison of the various capital charges for the various cases will provide the relative product costs.

In order to place all the cases under the same baseline, the "capital charges" should be referred to a unit of the products which are thought to be marketable.

Tacitly implied in the above conclusion is the requirement that the stream factor for all cases be the same, the necessary duplication and redundance, where required, of equipment will have to be provided to comply with it.

3. Time Frame Reference

This criteria will permit the elimination of the cases which, at the projected time frame, are not yet technologically perfected in a large enough scale as to be relied upon in a commercial venture.

Certain gray areas may be present when trying to apply this criteria and considerable subjective judgement may have to be exercised.

An example would be a Facility comprising the British Gas Corp./Lurgi slagging gasifiers.

This type of gasifier is not yet in commercial operation, although considerable testing for extended periods of time has been done on them. The tests were conducted with a 6' diameter gasifier, and the reports indicate satisfactory performance and yields.

One could say that at least for up to 6' and even maybe 8' diameter, it could be construed that the status is equivalent to a commercial status, although is not so in the strict sense.

4. The above listed criteria presuppose that all the cases considered are technically feasible for the processing of the type of coal selected and are environmentally acceptable.

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

PRODUCTS QUALITY

GASOLINE FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS WITH ESTIMATED FISCHER-TROPSCH GASOLINE PROPERTIES -

	Estimated F-T Unleaded Gasoline Properties	Specifications
Gravity, °API	67.2	
Octane Numbers		
Research	91	
Motor	83	82 min.
(Research + Motor)/2	87	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation °F		
IBP	86.5	
10%	108.9	158 max.
30%	137.3	
50%	186.0	170/250 min./max
70%	249.6	
90%	335.2	374 max.
EP	420.6	437 max.
V/L Ratio (=20), °F	@127.3	@140 max.
Sulfur, wt. %	Nil	0.10 max.
Composition, Vol. %		
Paraffins	60.5	
Olefins	19.9	20 max. (target)
Naphthenes	3.1	
Aromatics	16.5	
Molecular Weight	93.1	

SOURCE: D.O.E. FE-2447-9 Jan. 1978

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

PRODUCTS QUALITY

GASOLINE FROM F-T SYNTHESIS

EXPECTED BLEND COMPOSITION

Blending

Component	Wt. %
Mixed butanes	2.7
Light hydrocarbons	2.9
Alkylate	2.7
Light polymer gasoline	16.3
Hydrogenated heavy polymer gasoline	14.3
C_5/C_6 isomerate	25.1
Reformate	23.8
Hydrotreated gasifier naphtha	11.0
	100.0

Properties

Estimated properties and a comparison with unleaded gasoline specifications are presented elsewhere in this report.

This gasoline meets all specifications. Compared to typical present-day gasolines, it is slightly lower in gravity and contains fewer aromatics.

Heavy polymer gasoline hydrogenation has been added in order to meet the olefin target of 20 vol. % max. Without hydrogenation, the following are estimated:

(Research + Motor)/2 = 88.5
Olefin, vol. % = 28

Mobil and industry experience with high olefinic gasolines is limited. Consequently, marketing such a gasoline would require extensive testing and, perhaps, additive package reformulation.

SOURCE: D.O.E. FE-2447-9, Jan. 1978

PRODUCT QUALITY

DIESEL FUEL FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL DIESEL FUEL SPECIFICATIONS WITH ESTIMATED FISCHER-TROPSCH DIESEL FUEL PROPERTIES

	DIESEL FUEL-ASTM D975[1]			Estimated F-T Diesel Fuel Properties
	Grade No. 1-D	Grade No. 2-D	Other	
Gravity, ^{o}API	-	-		56
Cetane Number, min. (D613 or D976)	40	40	45[2]	50
Volatility				
90% min./max. ^{o}F (D86)	-/550	540/640[5]		520
Flash Point, ^{o}F min. (D93)	100	125		125
Cloud Point, ^{o}F (D2500)	(3)	(3)		10-20
Viscosity, Kin. @ 100^{o}F min./max. (D445)	1.4/2.5	2.0/4.3		2
Sulfur, wt. % max. (D129)	0.50	0.50	(4)	None
Corrosion, max. (D130)	No. 3	No. 3		No. 1
Carbon Residue on 10% Resid., max. (D524)	0.15	0.35		Trace

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(1) 1976 Annual Book of ASTM Standards.
(2) Various engine builders' requirements.
(3) Varies with marketing region; maps included in ASTM D975.
(4) Varies in individual states, some as low as 0.3%.
(5) 540/680 90% min./max. proposed but currently not accepted by industry.

SOURCE: D.O.E., FE-2447-9, January 1978.

PRODUCTS QUALITY

SNG EXPECTED COMPOSITION FROM F - T SYNTHESIS

Composition

Hydrogen	3.8
Methane	89.7
Ethene	1.0
Ethane	2.3
Propene	1.0
Propane	0.1
Carbon Dioxide	0.5
Inerts (N_2 + Ar)	1.6
	100.0

Other

Heat of Combustion (HHV)	1,003 Btu/SCF
Carbon monoxide (0.1% max.)	0.07%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

Index	Calculated	Preferable	Objectionable
Lifting, I_l	0.99	under 1.0	above 1.06
Flash-back, I_f	1.06	under 1.15	above 1.2
Yellow-Tip, I_y	0.97	above 1.0	under 0.8

The SNG, product is of satisfactory quality and is interchangable with pure methane. The olefin nature of the Fischer-Tropsch light gases, however, gives rise to a yellow-tip index value slightly below the preferable value. If the olefins prove troublesome in actual tests, an expensive hydrogenation step for olefin saturation could be added.

SOURCE: D.O.E. FW-2447-9, Jan. 1978

PRODUCTS QUALITY

HEAVY FUEL OIL FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL FUEL OIL AND GAS TURBINE FUEL OIL SPECIFICATIONS WITH ESTIMATED FISCHER-TROPSCH HEAVY FUEL OIL PROPERTIES

	Fuel Oil No. 5 (Light) ASTM D396*	Gas Turbine F.D. No. 5 - GT ASTM D2880*	Estimated F-T Hvy. Fuel Oil Properties
Gravity, °API	-	-	41
Volatility			
Flash Point, °F min. (D93)	130 or Legal	130 or Legal	150
Distillation	-	-	-
Ash, wt. % max. (D482)	0.10	0.03	Trace
Viscosity, SSU at 100°F, min. (D445)	150	45	Satisfactory
Sulfur, wt. % (D129)	Legal	-	None
Metals, ppm (wt.) max.			
Vanadium	-	2	0
Sodium + Potassium	-	5	Trace
Calcium	-	10	0
Lead	-	5	0

*1976 Annual Book of ASTM Standards.

SOURCE: D.O.E. FE-2447-9. Jan. 1978

PRODUCTS QUALITY

C_3LPG & C_4LPG FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL LPG SPECIFICATIONS WITH ESTIMATED FISCHER - TROPSCH LPG PROPERTIES-

	Commercial Propane ASTM D1835*	Est. MTG C_3 LPG Properties	Commercial Butane ASTM D1835*	Est. MTG C_4 LPG Properties
Vapor Pressure at 100°F, max., psig (D1267)	208	178	70	37
Volatile Residue				
Butane and Heavier, max. vol. % (D2163)	2.5	1	-	-
Pentane and Heavier, max. vol. % (D2163)	-	-	2.0	1
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

*1976 Annual Book of ASTM Standards.

Source: D.O.E., FE-24-2447-9, Jan. 1978

Job No.: B-28-8011
Date: November, 1981

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PRODUCTS QUALITY

ALCOHOLS EXPECTED COMPOSITION FROM F - T SYNTHESIS

Composition

Composition of the mixed alcohol product is:

Ethanol	59
I-propanol	14
N-propanol	14
2-butanol	4
I-butanol	1
N-butanol	5
2-pentanol	1
N-pentanol	1
C_6 plus	1
	100

The alcohol product is essentially free of acids, aldehydes, ketones and water.

Source, D.O.E. FE-2447-9, Jan. 1978

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PROCESS DESCRIPTION

INPUT-OUTPUT SUMMARIES

FLOWSHEETS

CASE I

LURGI (DRY ASH) GASIFIERS TO METHANOL TO GASOLINE

Process Description*

The coal to be fed to the gasifiers is crushed and graded between about 1/4" and 2" approximately.

The coal fines, -1/4", have been assumed to constitute 25% of the coal fed to the facility.

A portion of the coal fines is used as fuel for the generation of the captive steam and power, the remaining being used to generate power for export.

The graded coal is charged to the gasifiers via automatically operated coal lockhoppers. The bed of coal is gasified in countercurrent flow with the gasification agents; oxygen and steam. While traveling from the top to the bottom of the gasifiers, the coal is dried, devolatilized and gasified.

The steam and oxygen are introduced into the bottom of the gasifiers through a rotating grate. The ash is removed by this grate and discharged via lockhoppers.

The steam addition moderates the reaction temperature below the ash softening point.

The partial combustion of the coal with oxygen supplies the necessary heat for gasification. The raw gas exiting the gasifier is scrubbed and cooled with recyled liquor, thus removing most of the dust and the heavy tars.

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* Refer to the attached flowsheet.

The raw gas if further cooled in a waste heat boiler. The condensed liquors are sent to separators, where the aqueous and oil phases are separated. The oily phase, consisting of tars and oils is fed to the boilers for fuel.

The aqueous phase, containing phenols and ammonia is subjected to an extraction operation that removes the crude phenols, which in the present case are also used for fueling the boilers.

Ammonia containing water is stripped in the ammonia recovery section, and the water is sent to waste water treatment.

The raw gas will require some shifting in order to meet the H_2/CO ratio requirements for the methanol synthesis.

The shifted gas is processed in an acid gas removal plant (Rectisol) where essentially all of the sulfur compounds, H_2S and COS, the naphtha, HCN, CO_2 and water are removed by absorption in cold methanol. The naphthas are hydrotreated and blended into the gasoline product.

The methanol synthesis, effected by the Lurgi low pressure methanol process(1), reacts approximately 85% of the CO plus CO_2 into methanol. The crude methanol produced is a suitable feed stock for Mobil's Methanol to Gasoline process.

A slip stream of purge gas from methanol synthesis, which contains most of methane generated in the gasification step, unreacted CO, CO_2, H_2 and small amounts of C_2's, is treated in the H_2 recovery system for removal of the hydrogen required for the hydrotreating of the naphthas. The hydrogen is recovered from the feed gas by a pressure swing absorption operation, which

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(1) Process used for convenience in preparing this report.

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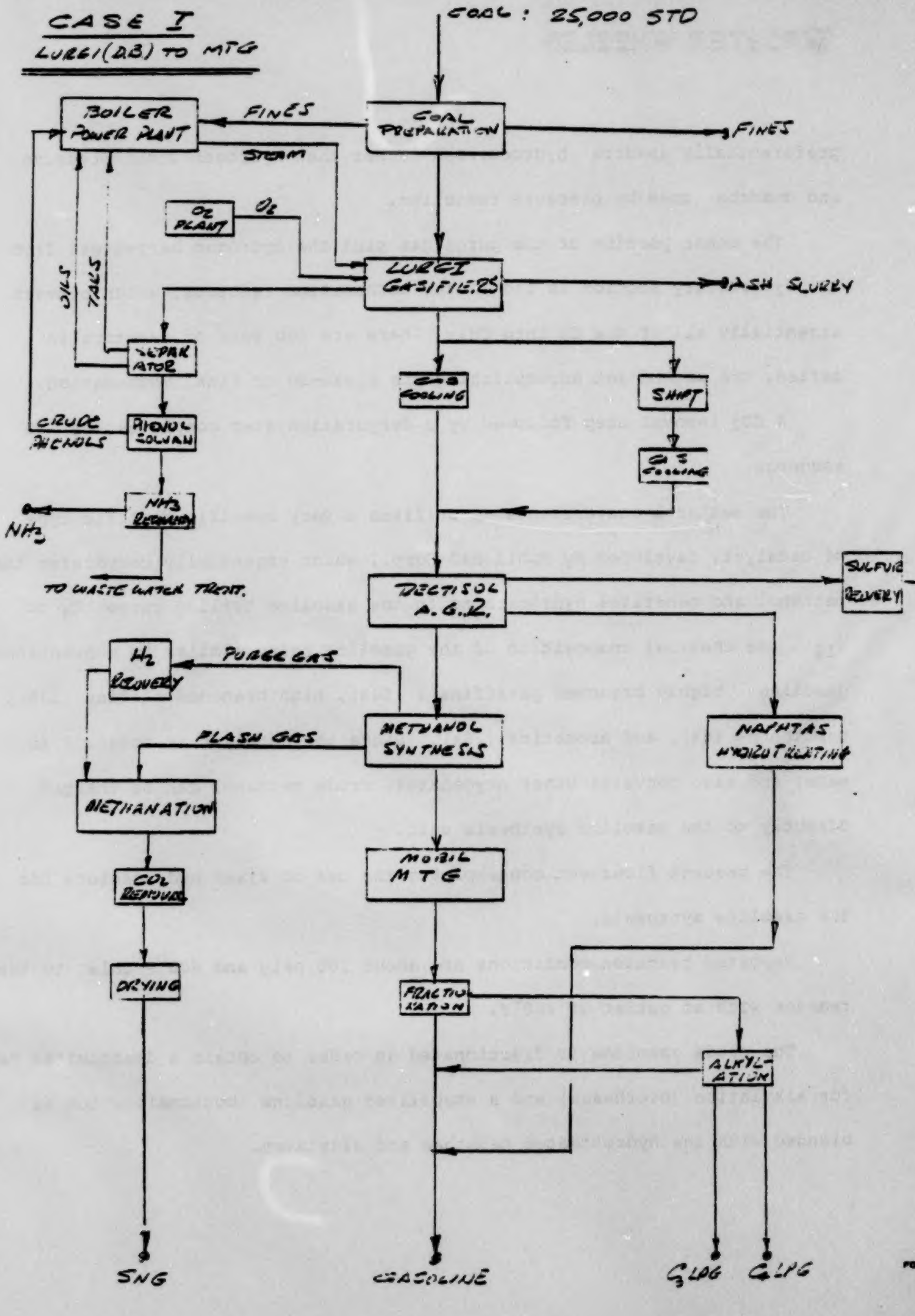
BY A G DATE 10/8/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE SCREENING OF APPLICABLE PROCESSES JOB NO. 8-28-3011

CASE I

LURGI (DB) TO MTG



preferentially absorbs hydrocarbons rather than hydrogen under pressure and desorbs them by pressure reduction.

The other portion of the purge gas plus the hydrogen barren gas from the H_2 recovery section is fed to the methanation reactors, which convert essentially all of the CO into CH_4. There are two sets of reactors in series, the second set accomplishing the clean-up or final methanation.

A CO_2 removal step followed by a dehydration step completes the SNG sequence.

The methanol conversion step utilizes a very specific zeolitic type of catalyst, developed by Mobil R&D Corp., which essentially dehydrates the methanol and generates hydrocarbons in the gasoline boiling range (C_4 to C_{10}), the chemical composition of the gasoline being similar to conventional gasoline: highly branched paraffins (50%), high branched olefins (13%), naphthenes (8%), and aromatics (28%). Since the catalyst is tolerant to water and also converts other oxygenates, crude methanol can be charged directly to the gasoline synthesis unit.

The present flowsheet contemplates the use of fixed bed reactors for the gasoline synthesis.

Reported reaction conditions are about 300 psig and 600°F inlet to the reactor with an outlet of 800°F.

The crude gasoline is fractionated in order to obtain a deethanized feed for alkylation (overheads) and a stabilized gasoline (bottoms), which is blended with the hydrotreated naphthas and alkylates.

The overhead from fractionation are treated with hydrofluoric acid to catalytically react butenes, and propenes with iso-butanes to produce higher molecular weight compounds, such as iso-octane.

C_3 and C_4 LPG are obtained as byproducts from the alkylation unit.

CASE I

INPUT - OUTPUT SUMMARY

INPUTS:

Coal:	HHV Btu/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUTS:

Product	Quantity	Btu/Unit	BTU/SD	%
SNG	92×10^6 SCFSD	1.000	9.20×10^{10}	40.67
Gasoline	15,070 Bbl/SD	5.1×10^6	7.68×10^{10}	33.96
C_3 LPG	1,059 Bbl/SD	3.8×10^6	4.02×10^9	1.77
C_4 LPG	1,500 Bbl/SD	4.2×10^6	6.30×10^9	2.78
Electricity	4.71×10^6 KWHR/SD	1×10^4	4.71×10^{10}	20.82
		TOTALS:	2.2622×10^{11}	100.00

F.O.E. (Fuel Oil Equivalent) @ 6 MMBTU/Bbl : 29,852 Bbl/SD

A. Overall Thermal Efficiency (Electrical Output Included): 66.6%

B. Thermal Efficiency (Electrical Output Excluded) : 52.8%

C. Coal Fines for Export in Case B: 3,475 TPSD

CASE IA

LURGI DRY BOTTOM GASIFIERS TO MEG WITH MAXIMIZING OF LIQUID PRODUCTS (SNG REFORMATE TO RECYCLE)

Process Description*

The flowsheet for this case is similar to Case I. An SNG reforming train has been substituted for the methanation unit.

The methanol synthesis purge gas is processed through the hydrogen recovery train, where the hydrogen is separated from the other components in the purge gas and a portion of it used for the hydrotreating of the naphthas. The remaining hydrogen, which is of about 98% purity with methane and CO_2 as its main impurities, is recycled to the syngas leaving the reforming plant.

The separation of hydrogen from the heavier components is achieved by the use of molecular sieves that adsorb the heavier components (CH_4, CO, CO_2, C_2+) under pressure, letting the hydrogen pass through. The adsorbed gases are desorbed by reducing the pressure, and the entire cycle is known as a pressure swing operation.

The hydrogen barren gas leaving the hydrogen recovery plant contains the CH_4 produced in the Lurgi gasifiers and some minor quantities of CO, CO_2 and C_2.

The gas is subjected to a partial oxidation step by reacting it with tonnage oxygen in the proper stoichiometric ratio to produce CO and H_2.

The Texaco and Shell partial oxidation processes, which are in commercial status, are examples of plants performing this operation.

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* Refer to attached flowsheet.

The required ratio of $H_2/(CO + CO_2)$ for the methanol or Fischer-Tropsch synthesis is adjusted by addition of the hydrogen recycle stream above mentioned.

The partial oxidation should be preferentially conducted at the pressure level of the synthesis reactors to economize in compression power.

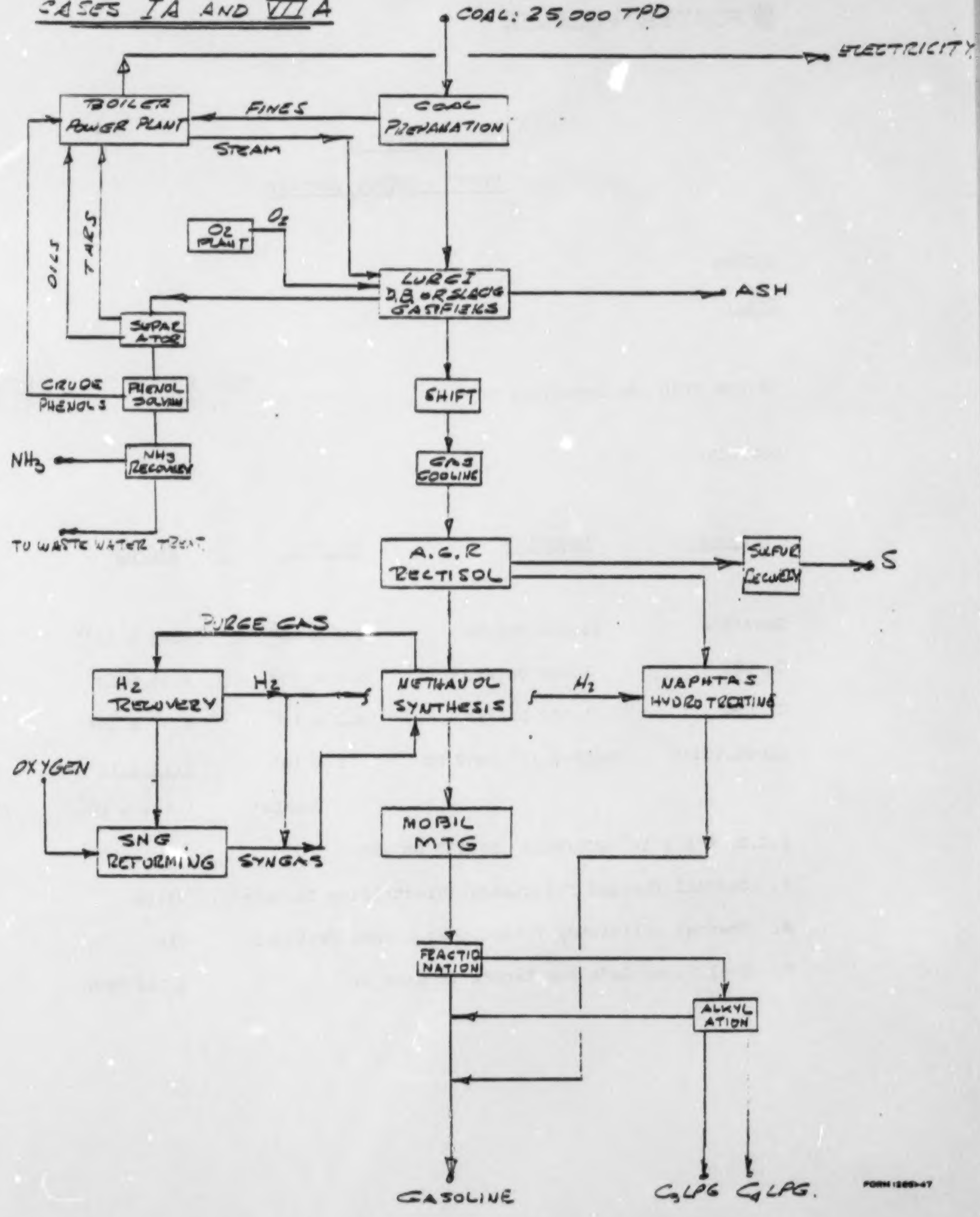
The reason being that the product gas $(CO + H_2)$ is about twice the volume of the feed gas $(CH_4 + O_2)$.

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N.J.

BY A. G DATE 10/27/81 SUBJECT SHEET NO. 1 OF 2

CHKD. BY DATE JOB NO. B-23-8011

CASES IA AND VIIA



FORM 1269-47

CASE IA

INPUT - OUTPUT SUMMARY

INPUTS

Coal:

	HHV, BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUTS:

Product	Quantity	Btu/Unit	Btu/SD	%
Gasoline	25,190 Bbl/SD	5.1×10^6	1.285×10^{10}	70.45
C_3 LPG	1.059 Bbl/SD	3.8×10^6	4.02×10^9	2.20
C_4 LPG	1,500 Bbl/SD	4.2×10^6	6.3×10^9	3.45
Electricity	4.36×10^6 KWHR/SD	1×10^4	4.36×10^{10}	23.90
		TOTALS:	1.824×10^{11}	100.00

F.O.E. @ 6×10^6 BTU/Bbl: 23,137 Bbl/SD

A. Overall Thermal Efficiency (Electricity Included): 53.8%

B. Thermal Efficiency (Electrical Output Excluded): 41%

C. Coal Fines Left for Export In Case B: 3,212 TPSD

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE II

LURGI DRY BOTTOMS TO FISCHER-TROPSCH SYNTHESIS

Process Description*

The coal preparation, gasification, gas cooling, gas shifting, and gas removal and tars, oils, phenols, and ammonia recovery are similar to the ones described in Case I.

The Fischer-Tropsch Synthesis here contemplated is based on the fast fluid bed (entrained recirculation) synthol synthesis reactors.

Iron based catalysts, promoted for improved selectivity and activity, and which are prepared and formulated for the end products desired are used.

The reported reactor operating conditions are approximately 600°F and 300 psig.

The reactions taking place being highly exothermic, one of the main problems in the reactors design is the heat removal mechanism. Internal tube-bundles with oil recirculation are reported to be used in conjunction with steam generators.

The $H_2/(2CO + 3CO_2)$ ratio seems to be critical for the obtention of the desired product slate.

The overall reactions occuring can be represented by the following simplified equations:

$$n\,CO + 2n\,H_2 \longrightarrow (-CH_2-)n + n\,H_2O$$

$$2n\,CO + n\,H_2 \longrightarrow (-CH_2-)n + n\,CO_2$$

$$n\,CO + 2n\,H_2 \longrightarrow H(-CH_2-)n\,OH + (n-1)\,H_2O$$

$$CO + H_2O \longrightarrow H_2 + CO_2$$

---

* Refer to attached flowsheet.

The products of the Synthol reactor include paraffins, mono-olefins, aromatics, alcohols, aldehydes, ketones, and fatty acids. Branching also occurs, predominantly as single-methyl branched structures.

The purified synthesis gas leaving the Rectisol plant is mixed with recycle gas and fed into the Synthol reactors. The gas flows upwards, after mixing with the circulating catalyst, the reaction heat being removed by the circulating oil in the tube-bundles immersed in the reactor. The reacted gas then enters a disengagement chamber where most of the entrained catalyst is separated from the gas by cyclones and recycled. The gas is first quenched by heat exchange with various streams and by recycled condensed oils, and finally water washed. The washed water contains the alcohols and acids.

There are four main streams leaving the F-T plant: a light oil fraction, a heavy oil fraction, an aqueous phase containing the alcohols and a gaseous purge stream which contains most of the methane, unreacted CO and H_2 and hydrocarbons.

The oil fractions are subsequently fractionated, the aqueous phase is treated in the alcohol recovery section and the purge stream is washed with heptane for hydrocarbon recovery. Most of the C_2's and heavier hydrocarbons in the purge stream are absorbed in cold heptane (approx. -40°F).

The loaded heptane is then reboiled and desorbed hydrocarbons are returned to the fractionation columns.

A slip stream of the methane containing purge gas leaving the heptane wash is sent to the hydrogen recovery section where most of its hydrogen is separated out by molecular sieves which operate in the pressure swing mode.

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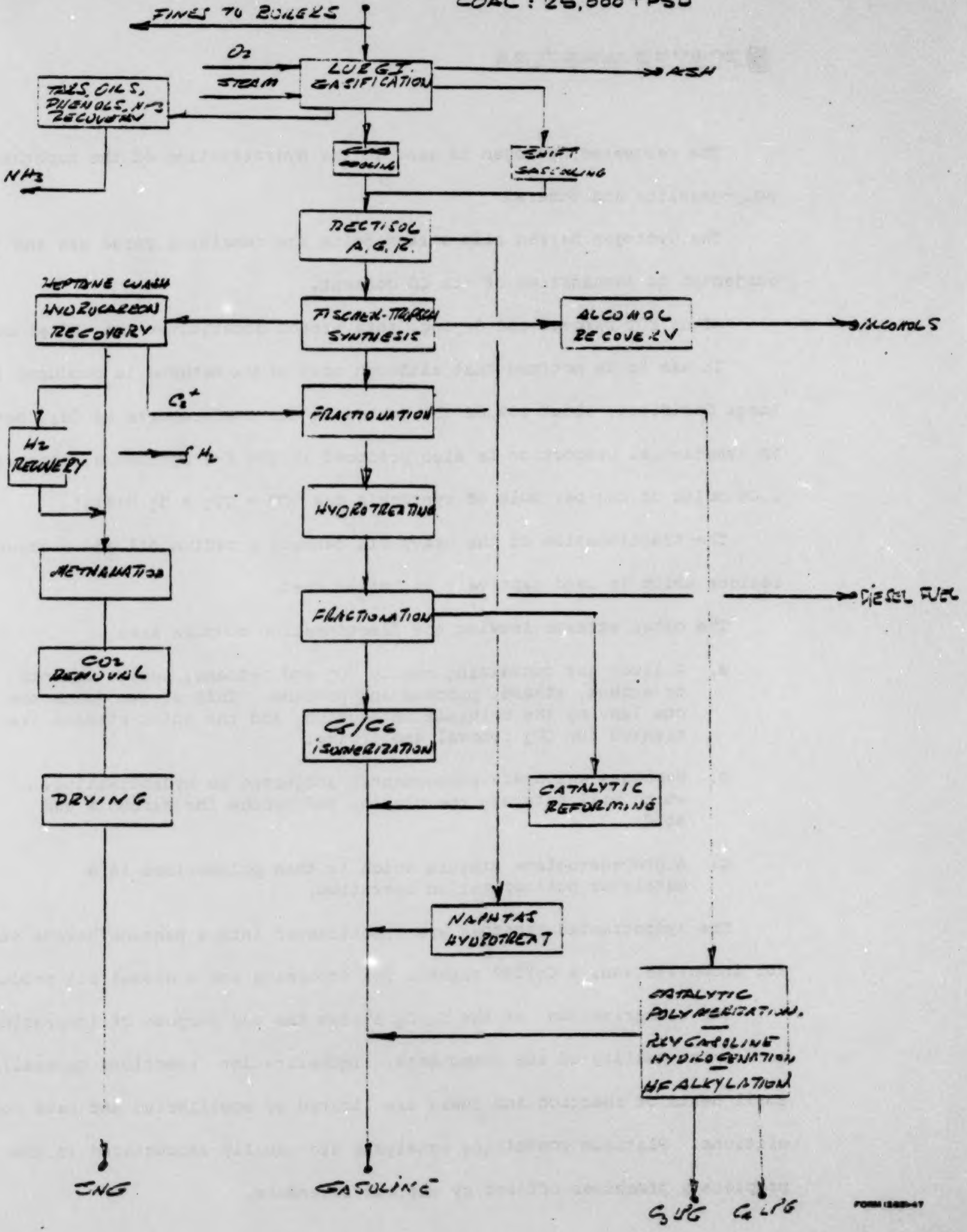
BY A. G. DATE 10/12/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE CASE II - LURGI (D.B.) → FISCHER-TROPSCH JOB NO. 3-28-8011

COAL : 25,000 TPSD



FORM 1383-47

The recovered hydrogen is used in the hydrotreating of the naphthas, poly-gasoline and others.

The hydrogen barren slip stream joins the remaining purge gas and is subjected to methanation of its CO content.

After CO_2 removal and drying, this stream constitutes the SNG product.

It has to be noticed that although most of the methane is produced in the Lurgi Gasifiers, about 18% of the carbon in the coal reports as CH_4, methane, in substantial proportion is also produced in the F-T synthesis, approximately 0.06 moles of CH_4 per mole of synthesis gas ($CO + CO_2 + H_2$ basis).

The fractionation of the heavy oil renders a medium oil and a vacuum residue which is used captively as boiler fuel.

The other streams leaving the fractionation complex are:

a. A light gas containing mostly CO_2 and methane, lesser amounts of ethene, ethane, propene and propane. This stream joins the one leaving the methanation section, and the joint streams are treated for CO_2 removal and drying.

b. Naphthas which are subsequently subjected to hydrotreating, which will saturate the olefins and reduce the alcohols and acids.

c. A propane-butene mixture which is then polimerized in a catalytic polimerization operation.

The hydrotreated naphthas are fractionated into a pentane/hexane stream for isomerization, a C_7/380 naphtha for reforming and a diesel oil product.

The isomerization of the C_5/C_6 stream has the purpose of increasing the anti-knock quality of the components. Isomerization reactions generally have small heats of reaction and these are limited by equilibrium and rate considerations. Platinum containing catalysts are usually encountered in the propietary processes offered by various licensors.

Catalytic reforming of the C_7/380 naphtha fraction is also practiced in this case for improvement in the octane rating. The most important reactions taking place in reforming are the ciclization of straight chain paraffins and dehydrogenation of same, isomerization of paraffins, dehydrogenation of naphtenes, and hydrocracking of paraffins and naphtenes.

The overall heat of reaction is endothermic, due mainly to the aromatization reactions, which are partially offset by the exothermic hydrocracking reactions.

Usually the heat of reaction is supplied by reheat of the streams between various staged reactors in series.

Catalysts employed are bimetallic catalysts such as platinum-rhenium, for example.

The overheads from the catalytic polimerization operation, which consist mainly of unreacted olefins, iso-butane, n-butane and propane, are alkylated with HF as a catalyst.

The alkylates are blended into the gasoline pool and the unreacted n-butanes and propane mixture are split into C_3 LPG and C_4 LPG.

CASE II

INPUTS - OUTPUT SUMMARY

INPUTS:

Coal:

	HHV, BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUTS:

Product	Quantity	HHV Btu/Unit	Btu/SD	%
SNG	104×10^6 SCFSD	1,000	1.04×10^{11}	53.44
Gasoline	8,915 Bbl/SD	5.1×10^6	4.55×10^{10}	23.38
Diesel Fuel	1,517 Bbl/SD	5.3×10^6	8.04×10^9	4.13
C_3 LPG	737 Bbl/SD	3.0×10^6	2.80×10^9	1.44
C_4 LPG	100 Bbl/SD	4.2×10^6	4.20×10^8	0.21
Alcohols	336,090 LB/SD	13,630	4.58×10^9	2.35
Electricity	2.93×10^6 KWHR/SD	1×10^4	2.93×10^{10}	15.05
		TOTALS:	1.946×10^{11}	100.00

F.O.E. @ 6 MMBTU/Bbl. : 27,550 Bbl/SD

A. Overall Thermal Efficiency (Electricity Included): 57.4%

B. Thermal Efficiency (Electricity Excluded): 48.7%

C. Coal Fines Left For Export In Case B : 2,162 TPSD

CASE IIA

LURGI DRY BOTTOM GASIFIERS TO FISCHER-TROPSCH SYNTHESIS WITH SNG REFORMING

Process Description*

The process flowsheet for this case is similar to Case II. An SNG reforming train has been substituted for the methanation unit.

The gas leaving the heptane wash, where most of its heavier components have been removed (C_2, C_3+'s) and returned to the fractionation section, joins the light ends overhead stream from the fractionation.

This stream is fed to the pressure swing hydrogen recovery plant, where molecular sieves adsorb nearly all of the components heavier then hydrogen.

The Fischer-Tropsch synthesis purge gas is processed through the hydrogen recovery train, where the hydrogen is separated from the other components in the purge gas and a portion of it used for the hydrotreating of the naphthas. The remaining hydrogen, which is of about 98% purity with methane and CO_2 as its main impurities, is recycled to the syngas leaving the SNG reforming plant.

The separation of hydrogen from the heavier components is achieved by the use of molecular sieves that absorb the heavier components (CH_4, CO, CO_2, C_2+) under pressure, letting the hydrogen pass through. The absorbed gases are desorbed by reducing the pressure, and the entire cycle is known as a pressure swing operation.

The hydrogen barren gas leaving the hydrogen recovery plant contains that CH_4 produced in the Lurgi gasifiers and some minor quantities of CO, CO_2 and C_2.

---

*Refer to attached flowsheet.

This gas is subjected to a partial oxidation step by reacting it with tonnage oxygen in the proper stoichiometric ratio to produce CO and H_2.

The Texaco and Shell partial oxidation processes, which are in commercial status, are examples of plants performing this operation.

The required ratio of $H_2/(CO + CO_2)$ for the methanol or Fischer-Tropsch synthesis is adjusted by addition of the hydrogen recycle stream above mentioned.

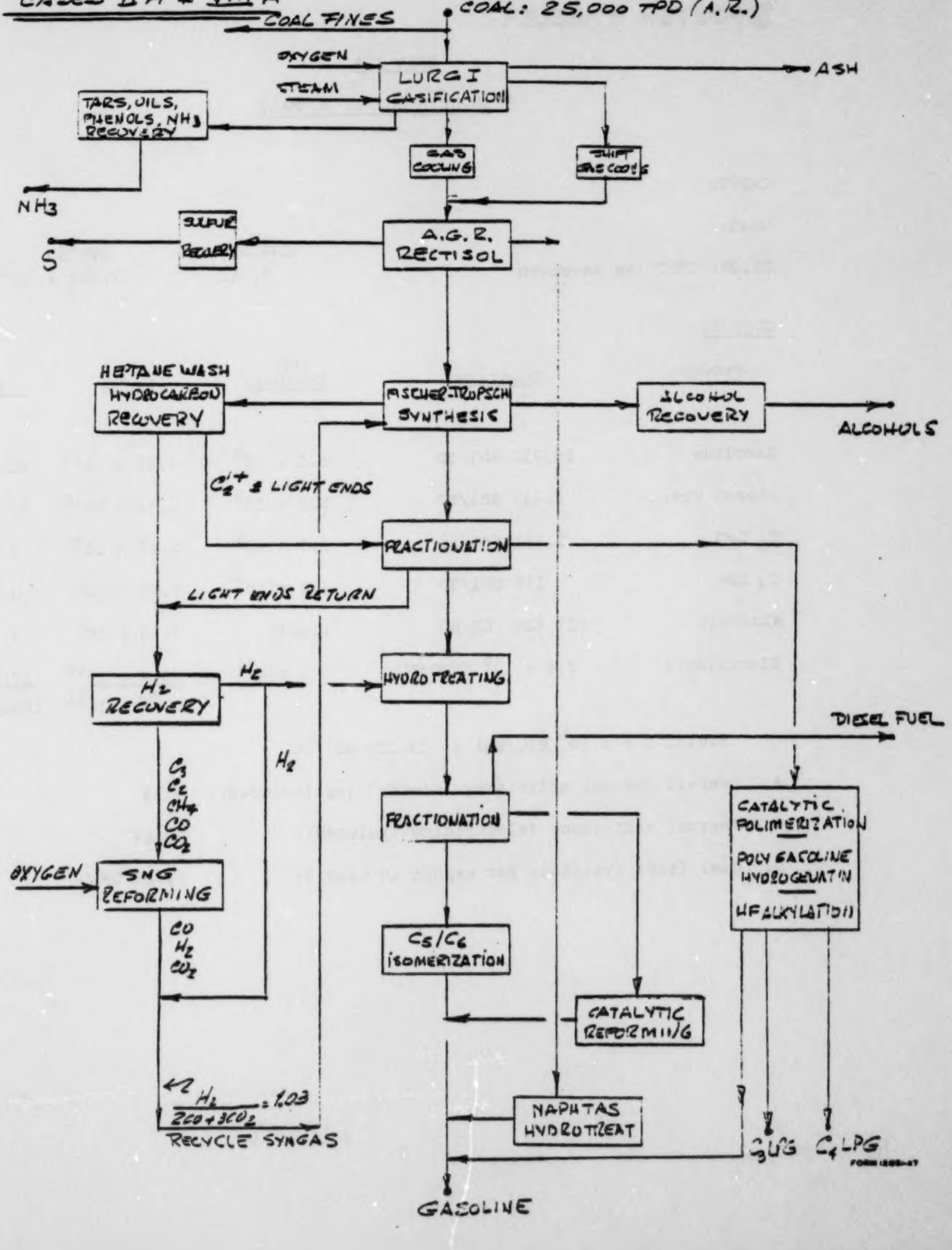
The partial oxidation should be preferentially conducted at the pressure level of the synthesis reactors to economize in compression power.

The reason being that the product gas ($CO + H_2$) is about twice the volume of the feed gas ($CH_4 + O_2$).

BY A G DATE 12/23/81 SUBJECT SHEET NO. 2 OF 2

CHKD. BY DATE JOB NO B-28-8011

CASES IIA & VIII A



FOSTER WHEELER

CASE IIA

INPUT - OUTPUT SUMMARY

INPUT:

Coal:

	HHV,BTU/LB	BTU/SD
25,000 TPSD (as Received	6,783	3.392×10^{11}

OUTPUTS:

Product	Quantity	HHV Btu/Unit	Btu/SD	%
Gasoline	16,712 Bbl/SD	5.1×10^6	8.52×10^{10}	61.39
Diesel Fuel	2,839 Bbl/SD	5.3×10^6	1.51×10^{10}	10.88
C_3 LPG	1,362 Bbl/SD	3.8×10^6	5.17×10^9	3.73
C_4 LPG	179 Bbl/SD	4.2×10^6	7.52×10^8	0.55
Alcohols	627,636 LB/SD	13,630	8.55×10^9	6.16
Electricity	2.4×10^6 KWHR/SD	1×10^4	2.4×10^{10}	17.29
			1.388×10^{11}	100.00

F.O.E. @ 6×10^6 BTU/Bbl : 19,128 Bbl/SD.

A. Overall thermal efficiency (electricity included): 41.%

B. Thermal efficiency (electricity excluded): 33.8%

C. Coal fines available for export in Case B: 1,767 TPSD

CASE III

WINKLER GASIFIER TO METHANOL TO GASOLINE

Process Description*

Coal is ground to -3/8" and dried to 8% moisture under nitrogen blanketing.

The ground coal is metered into the Winkler Gasifiers and a mixture of oxygen and steam is injected at various levels of the gasifier.

The operation of this type of gasifier involves a combination of fluidized bed and partial entrainment gasification. Coal at atmospheric pressure, or in this case at 60 psia, is fluidized by the mixture of oxygen and steam. The largest particles are gasified in the fluidized state while smaller particles in the feed or produced in the bed are gasified in suspension by injecting gasifying streams above the fluidized bed. A crude gas containing practically no tars or oils is thereby obtained. The freeboard in these gasifiers is about 5 to 10 times the height of the fluidized bed. The gasification temperature is kept below the ash fusion temperature by adjusting the oxygen-steam ratio.

The usually encountered gasification temperatures range between 1500°F to 1900°F, depending on the type of coal fed, and its reactivity.

The carbon utilization in these gasifiers is mainly controlled by the residence time of the particles and the reactivity of the coal.

About 70% of the total char generated will be entrained with the raw gas, the remaining leaving through the bottom of the gasifier.

*Refer to attached flowsheet.

The flow of gasifying agents injected over the top of the bed (freeboard) has to be accurately metered, if not enough it will result in non-gasified carbon in the ash leaving the top of the gasifiers, if too much it will result in a higher CO_2 containing raw gas, with a loss of potential product gas.

The carbon utilization assumed for the present case was taken at 84%.

The unreacted carbon leaving with the top and bottom of the ash, is returned to the Boiler Plant where it is mixed with ground lignite and fueled.

The raw gas passes through waste heat boilers, fitted with a radiant and convection sections, the radiant sections decreasing the temperature of the raw gas well below the ash softening point, in order to avoid fouling of the tubes in the convection section.

High pressure steam for power and process usage is generated in the waste heat boilers.

Cyclones following the heat recovery train remove most of the particulates in the raw gas, which is then subjected to a final clean-up in high impact venturi scrubbers,where the dust content of the gas is reduced to about 0.01 grain/SCF dry gas (approx. 25 ppm by weight).

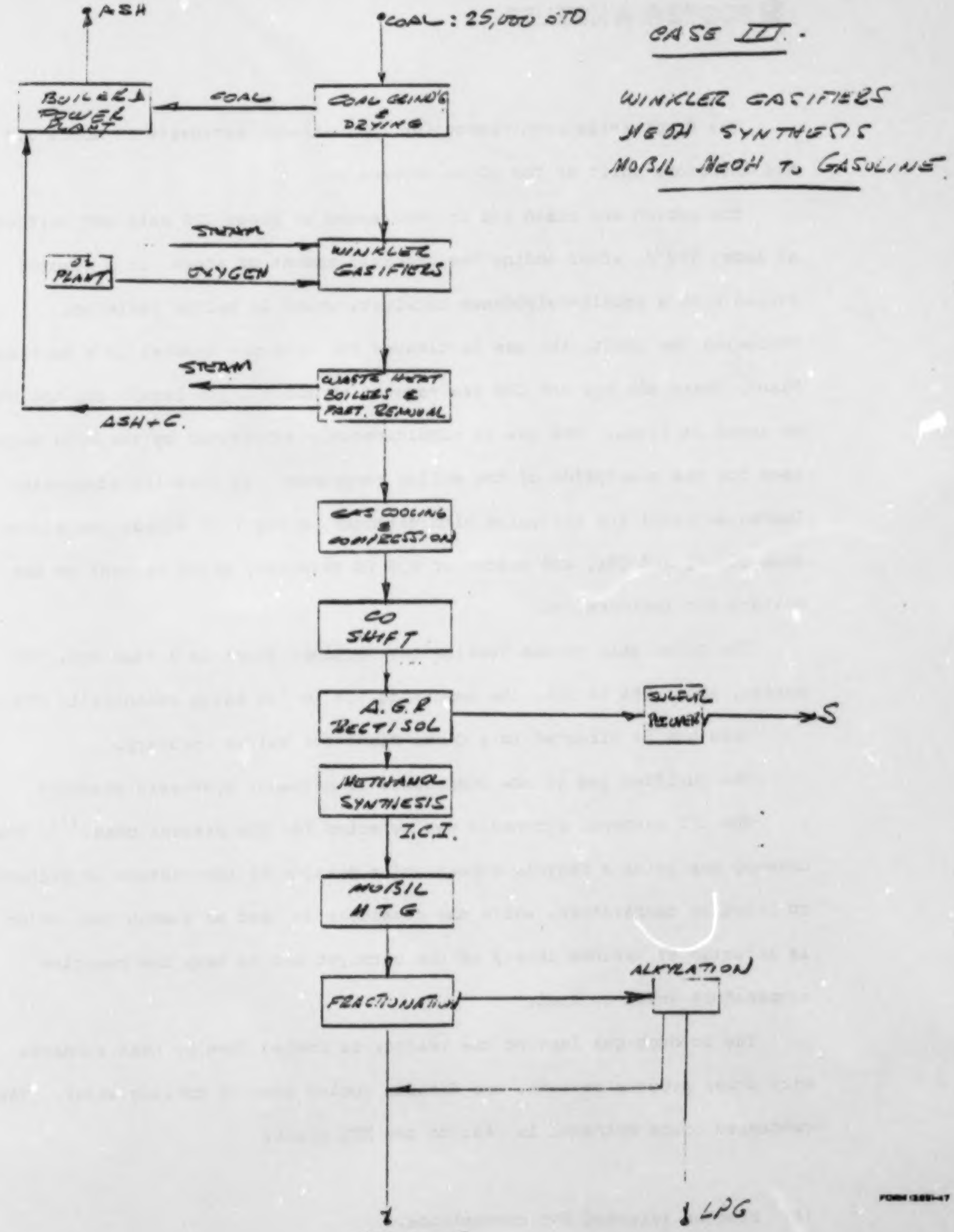
The cooled gas contains CO and H_2 in an approximate ratio of one.

BY G G DATE 10/14/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF.

CHKD. BY DATE CASE III - WINKLER TO MEOH TO GASOLINE JOB NO. B-28-8011

CASE III -

WINKLER GASIFIERS
MEOH SYNTHESIS
MOBIL MEOH TO GASOLINE



The H_2/CO ratio requirement for the methanol synthesis is about 2.4 and therefore shift of the CO is necessary.

The cooled and clean gas is compressed to about 700 psig and shifted at about 700°F, after adding the required amount of steam, in a reactor filled with a cobalt-molybdenum catalyst, which is sulfur resistant. Following the shift, the gas is treated for acid gas removal in a Rectisol Plant, where the H_2S and COS are removed to about 1 ppm level, and the CO_2 to about 4% level. The gas is simultaneously dehydrated by the cold methanol used for the absorption of the sulfur compounds. By flashing stage-wise the loaded methanol and stripping with nitrogen, a CO_2 rich stream containing some CO, H_2 and CH_4, and traces of H_2S is obtained, which is sent to the boilers for incineration.

The other main stream leaving the Rectisol plant is a rich H_2S, COS stream, about 25% to 30%, the remaining 70% to 75% being essentially CO_2.

This gas is directed to a Claus Plant for sulfur recovery.

The purified gas is now compressed to methanol synthesis pressure.

The ICI methanol synthesis was selected for the present case.(1) The make-up gas joins a recycle stream and a portion of the mixture is preheated to reaction temperature, while the remainder is used as quench gas, which is injected at various levels of the catalyst bed to keep the reaction temperature under control.

The product gas leaving the reactor is cooled down by heat exchange with other process streams, and finally cooled down by cooling water. The condensed crude methanol is sent to the MTG plant.

---

(1) Process selected for convenience.

Most of the non-condensed portion of the product gas is recycled to the front-end of the reactors and the remainder constitutes the purge stream for the inerts that accumulate in the system, such as methane and nitrogen.

The methanol to gasoline process, developed by Mobil Research and Development Corp., utilizes a very specific zeolitic type catalyst, which essentially dehydrates the methanol and produces hydrocarbons in the gasoline boiling range (C_4 to C_{10}).

The gasoline produced in this process has a chemical composition similar to conventional gasoline. The catalyst is water tolerant and also converts other oxgenates which make it suitable for charging with crude methanol.

Reaction reported conditions are about 300 psig and 600°F at the inlet of reactors and about 800°F at the outlet.

The crude gasoline from the MTG Synthesis is fractionated into an overhead suitable for aklylation and a stabilized bottoms gasoline.

The overhead comprising mainly iso-butene, iso-butane, propene, propane, is alkylated with HF as a catalyst to obtain higher iso-paraffins such as iso-octane for example, that are blended with the gasoline pool.

The unreacted propane and some butane constitute the LPG product.

Job No.: B-28-8011
Date: November 1981

FOSTER WHEELER

CASE III

INPUT - OUTPUT SUMMARY

INPUTS:

Coal:

	HHV,BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUTS

Product	Quantity	HHV BTU/Unit	BTU/SD	%
Gasoline	24,067 Bbl/SD	5.1×10^6	1.227×10^{11}	92.78
LPG	2,502 Bbl/SD	3.8×10^6	9.51×10^9	7.22
			1.322×10^{11}	100.00

F.O.E. @ 6 MMBTU/Bbl: 22,033 Bbl/SD

Overall Thermal Efficiency: 39%

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE IV

WINKLER GASIFIER TO FISCHER-TROPSCH TO GASOLINE

Process Description*

Coal is ground to -3/8" and dried to 8% moisture under nitrogen blanketing.

The ground coal is metered into the Winkler Gasifiers and a mixture of oxygen and steam is injected at various levels of the gasifier.

The operation of this type of gasifier involves a combination of fluidized bed and partial entrainment gasification. Coal at atmospheric pressure, or in this case at 60 psia, is fluidized by the mixture of oxygen and steam. The largest particles are gasified in the fluidized state while smaller particles in the feed or produced in the bed are gasified in suspension by injecting gasifying streams above the fluidized bed. A crude gas containing practically no tars or oils is thereby obtained. A freebaord in these gasifiers is about 5 to 10 times the height of the fluidized bed. The gasification temperature is kept below the ash fusion temperature by adjusting the oxygen-steam ration.

The usually encountered gasification temperatures range between 1500°F to 1900°F, depending on the type of coal fed and its reactivity.

The carbon utilization in these gasifiers is mainly controlled by the residence time of the particles and the reactivity of the coal.

About 70% of the total char generated will be entrained with the raw gas, the remaining leaving through the bottom of the gasifier.

---

* Refer to attached flowsheet.

The flow of gasifying agents injected over the top of the bed (freeboard) has to be accurately metered, if not enough it will result in non-gasified carbon in the ash leaving the top of the gasifiers, if too much it will result in a higher CO_2 containing raw gas, with a loss of potential product gas.

The carbon utilization assumed for the present case was taken at 84%.

This unreacted carbon leaving with the top and bottom of the ash, is returned to the Boiler Plant where it is mixed with ground lignite and fueled.

The raw gas passes through waste heat boilers, fitted with a radiant and convection sections, the radiant sections decreasing the temperature of the raw gas well below the ash softening point, in order to avoid fouling of the tubes in the convection sections.

High pressure steam for power and process usage is generated in the waste heat boilers.

Cyclones following the heat recovery train remove most of the particulates in the raw gas, which is then subjected to a final clean-up in high impact venturi scrubbers, where the dust content of the gas is reduced to about 0.01 grain/SCF dry gas (approx. 25 ppm by weight).

The cooled gas contains CO and H_2 in an approximate ratio of one.

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N.J.

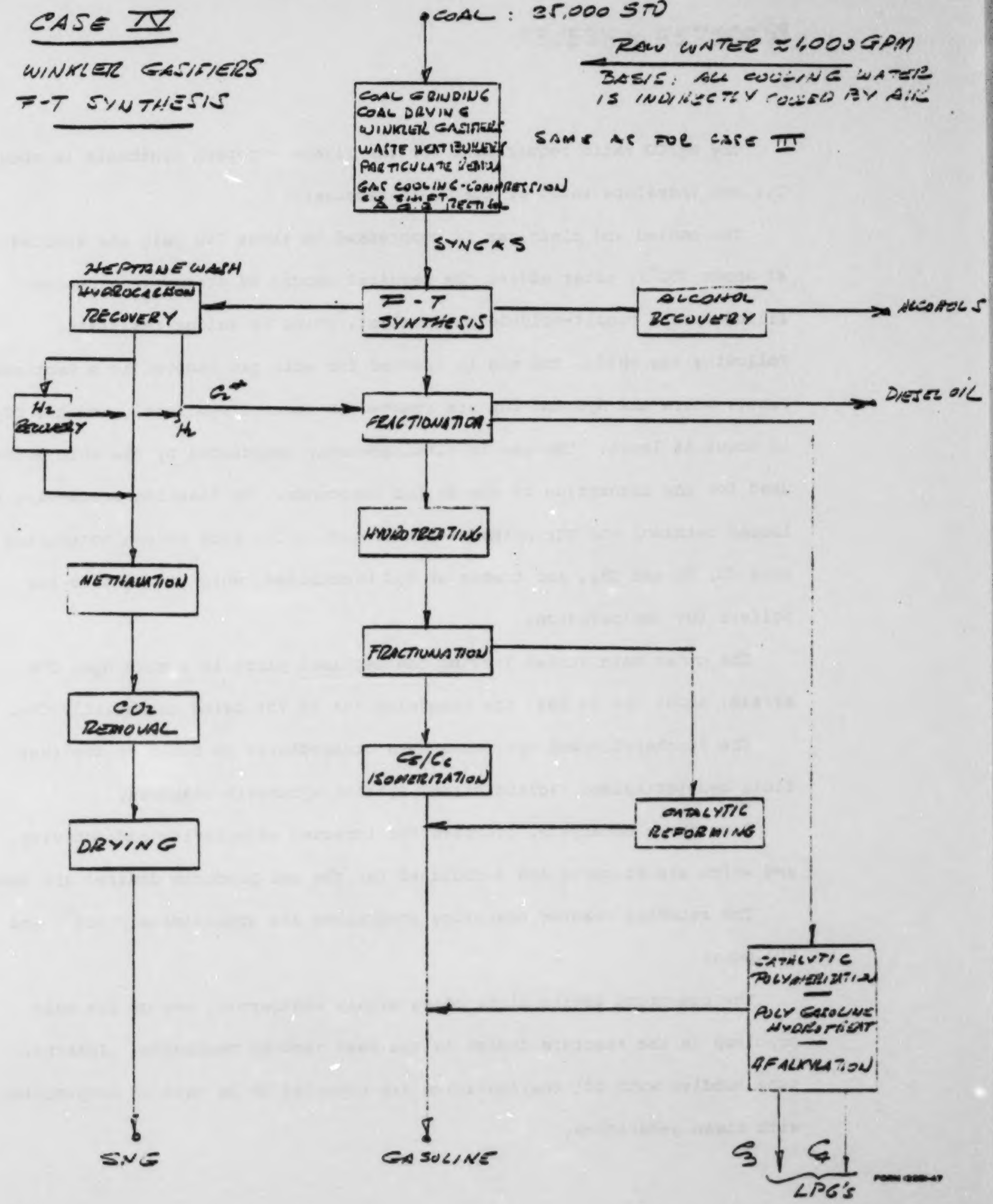
BY G G DATE 10/16/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE CASE IV - WINKLER TO FISCHER-TROPSH JOB NO 13-28-8011

CASE IV

WINKLER GASIFIERS

F-T SYNTHESIS



The H_2/CO ratio requirement for the Fischer-Tropsch synthesis is about 2.4 and therefore shift of the CO is necessary.

The cooled and clean gas is compressed to about 700 psig and shifted at about 700°F, after adding the required amount of steam, in a reactor filled with a cobalt-molybdenum catalyst, which is sulfur resistant. Following the shift, the gas is treated for acid gas removal in a Rectisol Plant, where the H_2S and COS are removed to about 1 ppm level, and the CO_2 to about 4% level. The gas is simultaneously dehydrated by the cold methanol used for the absorption of the sulfur compounds. By flashing stage-wise the loaded methanol and stripping with nitrogen, a CO_2 rich stream containing some CO, H_2 and CH_4, and traces of H_2S is obtained, which is sent to the boilers for incineration.

The other main stream leaving the Rectisol plant is a rich H_2S, COS stream, about 25% to 30%, the remaining 70% to 75% being essentially CO_2.

The Fischer-Tropsch Synthesis here contemplated is based on the fast fluid bed (entrained recirculation) Synthol synthesis reactors.

Iron based catalysts, promoted for improved selectivity and activity, and which are prepared and formulated for the end products desired are used.

The reported reactor operating conditions are approximately 600°F and 300 psig.

The reactions taking place being highly exothermic, one of the main problems in the reactors design is the heat removal mechanism. Internal tube bundles with oil recirculation are reported to be used in conjunction with steam generators.

The $H_2/(2CO + 3CO_2)$ ratio seems to be critical for the obtention of the desired product slate.

The overall reactions occuring can be represented by the following simplified equations:

$$n\,CO + 2n\,H_2 \longrightarrow (-CH_2-)n + n\,H_2O$$
$$2n\,CO + n\,H_2 \longrightarrow (-CH_2-)n + n\,CO_2$$
$$n\,CO + 2n\,H_2 \longrightarrow H(-CH_2-)n\,OH + (n-1)\,H_2O$$
$$CO + H_2O \longrightarrow H_2 + CO_2$$

The products of Synthol reactor include paraffins, mono-olefins, aromatics, alcohols, aldehydes, ketones, and fatty acids. Branching also occurs, predominantly as single-methyl branched structures.

The purified synthesis gas leaving the Rectisol plant is mixed with recycle gas and fed into the Synthol reactors. The gas flows upwards, after mixing with the circulating catalyst, the reaction heat being removed by the circulating oil in the tube-bundles immersed in the reactor. The reacted gas then enters a disengagement chamber where most of the entrained catalyst is separated from the gas by cyclones and recycled. The gas is first quenched by heat exchange with various streams and by recycled condensed oils, and finally water washed. The wash water contains the alcohols and acids.

There are four main streams leaving the F-T plant: a light oil fraction, a heavy oil fraction, an aqueous phase containing the alcohols and a gaseous purge stream which contains most of the methane, unreacted CO and H_2 and hydrocarbons.

The oil fractions are subsequently fractionated, the aqueous phase is treated in the alcohol recovery section and the purge stream is washed with heptane for hydrocarbon recovery. Most of the C_2's are heavier hydrocarbons in the purge stream are absorbed in cold heptane (approx. $-40^{\circ}F$).

The loaded heptane is then reboiled and desorbed hydrocarbons are returned to the fractionation columns.

A slip stream of the methane containing purge gas leaving the heptane wash is sent to the hydrogen recovery section where most of its hydrogen is separated out by molecular sieves which operate in the pressure swing mode.

The recovered hydrogen is used in the hydrotreating of the naphthas, poly-gasoline and others.

The hydrogen barren slip stream joins the remaining purge gas and is subjected to methanation of its CO content.

After CO_2 removal and drying, this stream constitutes the SNG product.

It has to be noticed that although some of the methane is produced in the Winkler Gasifiers, about 3% of the carbon in the coal reports as CH_4, methane in substantial proportion is also produced in the F-T synthesis, approximately 0.06 moles of CH_4 per mole of synthesis gas ($CO + CO_2 + H_2$ basis.)

The fractionation of the heavy oil renders a medium oil and a vacuum residue which is used captively as boiler fuel.

The other streams leaving the fractionation complex are:

a. A light gas containing mostly CO_2 and methane, lesser amounts of ethene, ethane, propene and propane. This stream joins the one leaving the methanation section, and the joint streams are treated for CO_2 removal and drying.

b. Naphthas which are subsequently subjected to hydrotreating, which will saturate the olefins and reduce the alcohols and acids.

c. A propane-butene mixture which is then polimerized in a catalytic polimerization operation.

The hydrotreated naphthas are fractionated into a pentane/hexane stream for isomerization, a C_7/380 naphtha for reforming and a diesel oil product.

The isomerization of the C_5/C_6 stream has the purpose of increasing the anti-knock quality of the components. Isomerization reactions generally have small heats of reaction and these are limited by equilibrium and rate considerations. Platinum containing catalysts are ususally encountered in the propietary processes offered by various licensors.

Catalytic reforming of the C_7/380 naphtha fraction is also practiced in this case for improvement in the octane rating. The most important reactions taking place in reforming are the ciclization of straight chain paraffins and dehydrogenation of same, isomerization of paraffins, dehydrogenation of naphthenes, and hydrocracking of paraffins and naphtenes.

The overall heat of reaction is endothermic, due mainly to the aromatization reactions, which are partially offset by the exothermic hydrocracking reactions.

Usually the heat of reaction is supplied by reheat of the streams between various staged reactors in series.

Catalysts employed are bimetallic catalysts such as platinum-rhenium, for example.

The overheads from the catalytic polimerization operation, which consist mainly of unreacted olefins, iso-butane, n-butane and propane, are alkylated with HF as catalyst.

The alkylates are blended into the gasoline pool and the unreacted n-butanes and propane mixture are split into C_3 LPG and C_4 LPG.

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE IV

INPUT - OUTPUT SUMMARY

INPUT:

Coal:	HHV,BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUT:

Product	Quantity	HHV BTU/Unit	BTU/SD	%
SNG	52.2×10^6 SCFD	1,000	5.22×10^{10}	40.56
Gasoline	10,786 Bbl/SD	5.1×10^6	5.50×10^{10}	42.74
Diesel Fuel	2,060 Bbl/SD	5.3×10^6	1.09×10^{10}	8.47
C_3 LPG	1,000 Bbl/SD	3.8×10^6	3.80×10^9	2.95
C_4 LPG	136 Bbl/SD	4.2×10^6	5.70×10^8	0.45
Alcohols	456,450 LB/SD	13,630	6.22×10^9	4.83
			1.287×10^{11}	100.00

F.O.E. @ 6MMTBTU/Bbl: 21,450 Bbl/SD

Overall Thermal Efficiency: 38%

Job No.: B-28-8011
Date: November, 1981

CASE V

KOPPERS-TOTZEK GASIFIER TO METHANOL TO GASOLINE

Process Description*

Coal is ground to about 90% through 200 mesh and dried to approximately 8% moisture.

The ground coal is transferred via a screw feeder at a controlled ratio to a mixing head, where it is mixed and entrained by a mixture of steam and oxygen into the gasifier.

The velocity of the coal particles through the nozzle should be kept higher than the flame propagation velocity to avoid flashbacks.

The gasifier typically has 2 or 4 heads facing each other. It is built with a double shell and is usually lined with refractory. Low pressure steam generated in the shell annulus keeps the temperature of the inner wall below the ash slagging temperature, thereby allowing the build-up of a layer of frozen slag that provided additional protection.

It has been reported that satisfactory operation has been achieved with high ash lignites without any refractory lining.

The burners are positioned directly opposite, which improves turbulence. If some coal particles pass unburned through one jet they will be caught in the opposite one and complete their combustion.

The temperatures at the flame are in the range of 3000°F to 3500°F and most of the carbon in the coal is reacted in the flame front.

---

* Refer to attached flowsheet.

There is very little methane formation and no tars or oils present in the raw gas. Approximately 50% of the ash forms molten slag that flows to the bottom of the gasifier where it is quenched with water; the remainder leaves with the raw gas.

The raw gas is partially quenched with water at the exit of the gasifier and then is cooled by a waste heat boiler, generating high pressure steam. Carbon utilization for lignite coals is in order of 98%. Gas leaving the waste heat boilers is washed and cooled with a circulating stream of water. Most of the flyash is removed here. Additional flyash is removed in the disintegrators and final gas clean-up is usually done with electrostatic precipitators.

The operation of the gasifier is at essentially atmospheric pressure, about 5 to 6 psig.

The cooled and cleaned gas is stored in gas-holders, from where it is transferred to the compression plant to bring it to shift pressure.

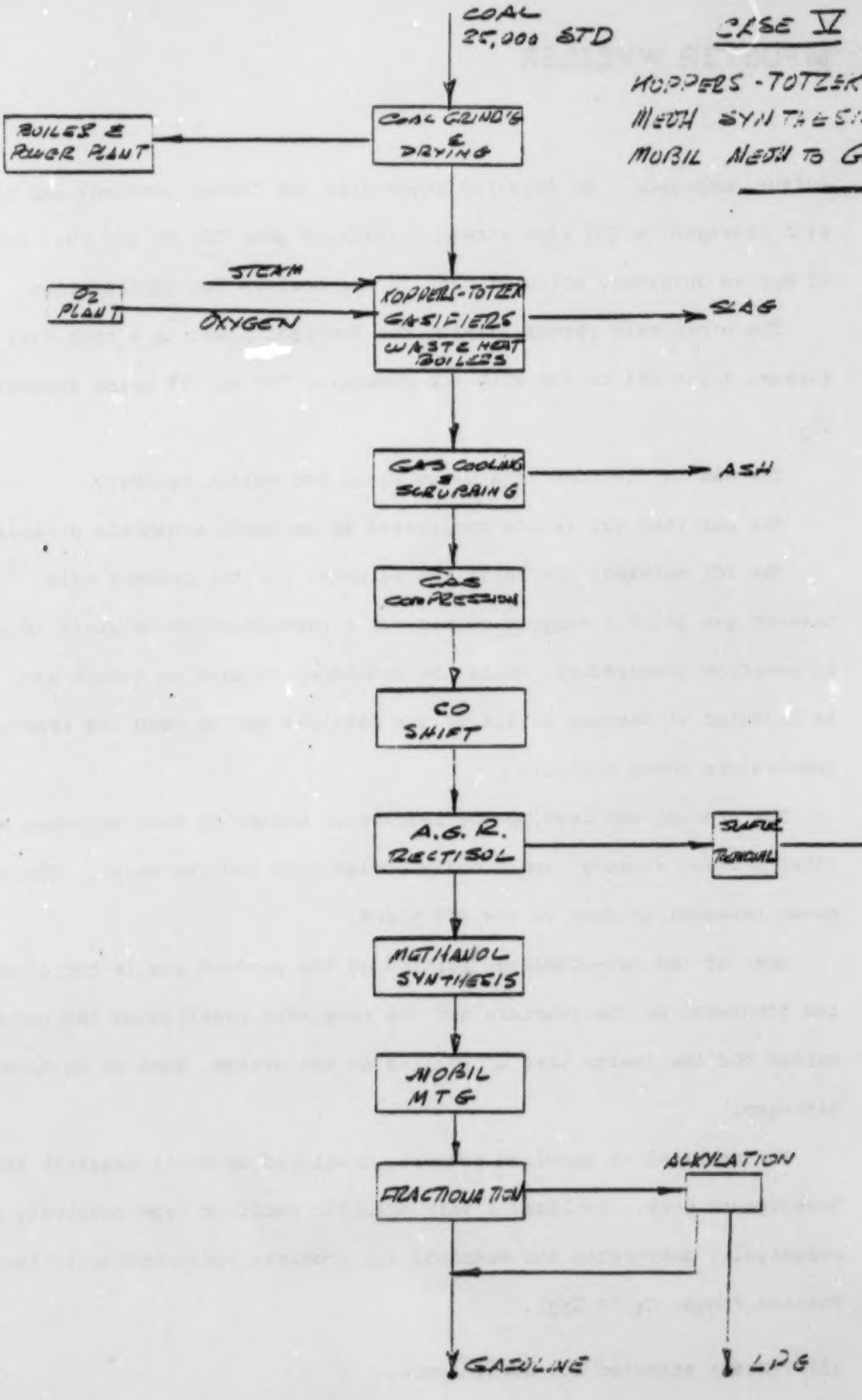
The H_2/CO ratio of this gas is about 2: 1.

The H_2/CO requirement for the methanol synthesis is about 2.4:1 and therefore a partially shifting of the CO is necessary.

The cooled and clean gas is compressed to about 700 psig and after adding the required amount of steam is shifted at about 700°F, in a reactor filled with a sulfur resistant cobalt-molybdunum catalyst. Following the shift, the gas is treated for acid gas removal in a Rectisol Plant, where the H_2S and COS are removed to about 1 ppm, and the CO_2 to about 4% level. The gas is simultaneously dehydrated by the cold methanol used for absorption of the

BY G.G. DATE 10/20/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF
CHKD. BY DATE CASE V - K-T TO MTG. JOB NO. 3-28-8011

CASE V
KOPPERS-TOTZEK GASIFIERS
METH SYNTHESIS
MOBIL METH TO GASOLINE



sulfur compounds. By flashing stage-wise the loaded methanol and stripping with nitrogen, a CO_2 rich stream containing some CO, H_2 and CH_4, and traces of H_2S is obtained, which is sent to the boilers for incineration.

The other main stream leaving the Rectisol plant is a rich H_2S, COS stream, about 25% to 30% with the remaining 70% to 75% being essentially CO_2.

The gas is directed to a Claus Plant for sulfur recovery.

The purified gas is now compressed to methanol synthesis pressure.

The ICI methanol synthesis was selected for the present case.(1) The make-up gas joins a recycle stream and a portion of the mixture is preheated to reaction temperature, while the remainder is used as quench gas, which is injected at various levels of the catalyst bed to keep the reaction temperature under control.

The product gas leaving the reactor is cooled by heat exchange with other process streams, and finally cooled with cooling water. The condensed crude methanol is sent to the MTG plant.

Most of the non-condensed portion of the product gas is recycled to the front-end of the reactors and the remainder constitutes the purge stream for the inerts that accumulate in the system, such as methane and nitrogen.

The methanol to gasoline process, developed by Mobil Research and Development Corp., utilizes a very specific zeolitic type catalyst, which essentially dehydrates the methanol and produces hydrocarbons in the gasoline boiling range (C_4 to C_{10}).

(1) Process selected for convenience.

The gasoline produced in this process has a chemical composition similar to conventional gasoline. The catalyst is water tolerant and also converts other oxgenates which make it suitable for charging with crude methanol.

Reaction reported conditions are about 300 psig and 600°F at the inlet of reactors and about 800°F at the outlet.

The crude gasoline from the MTG Synthesis is fractionated into an overhead, suitable for alkylation, and stabilized bottoms gasoline.

The overhead comprising mainly iso-butene, iso-butane, propene, propane, is alkylated with HF as a catalyst to obtain higher iso-paraffins such as iso-octane for example, that are blended with the gasoline pool.

The unreacted propane and some butane constitute the LPG product.

The Koppers-Totzek with the Fischer-Tropsch synthesis will provide product yields similar to the ones found for the Winkler - F-T case (Case IV), and it was considered not justifiable for the level of effort allowed for this report to develop these data.

CASE V

INPUT - OUTPUT SUMMARY

INPUT:

Coal:

	HHV,BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUT:

Product	Quantity	HHV BTU/Unit	BTU/SD	%
Gasoline	23,255 Bbl/SD	5.1×10^6	1.186×10^{11}	92.78
LPG	1,821 Bbl/SD	3.8×10^6	6.92×10^{11}	7.22
			1.22×10^{11}	100.00

F.O.E. @ 6 MMBTU/Bbl: 20,917 Bbl/SD

Overall Thermal Efficiency: 37%

CASE VI

TEXACO GASIFIER TO METHANOL TO GASOLINE

Process Description*

The coal is ground and slurried in water and fed under pressure through a special type of burner to the gasifier.

The size of the ground coal, its pretreatment, if any required, and the slurrying operation are propietary information of the Texaco Development Corp. Same applies to the burners and other parts of the Texaco Gasifiers.

The present case assumes a slurry concentration of 43%.**

The gasification is of the entrained downflow mode. The coal slurry, oxygen and steam are fed to the top of the gasifiers, and the flame front travels downwards.

Temperatures range in the order ot 2000°F to 2800°F. The hot gas is water quenched and the slag is removed from the gasifier bottom by a lock-hopper system.

The raw gas is further cooled in waste heat boilers, scrubbed in venturi contactors and sent to the shift section.

From here the flowsheet is identical to Case V and need no further description, except to note that the Texaco Gasifiers operate under pressures ranging up to 1200 psig, and consequently the gas compression section shown for Case V is not required in the present case.

---

* The flowsheet is similar to the Koppers-Totzek case, by substituting Texaco for K-T in the attached flowsheet.

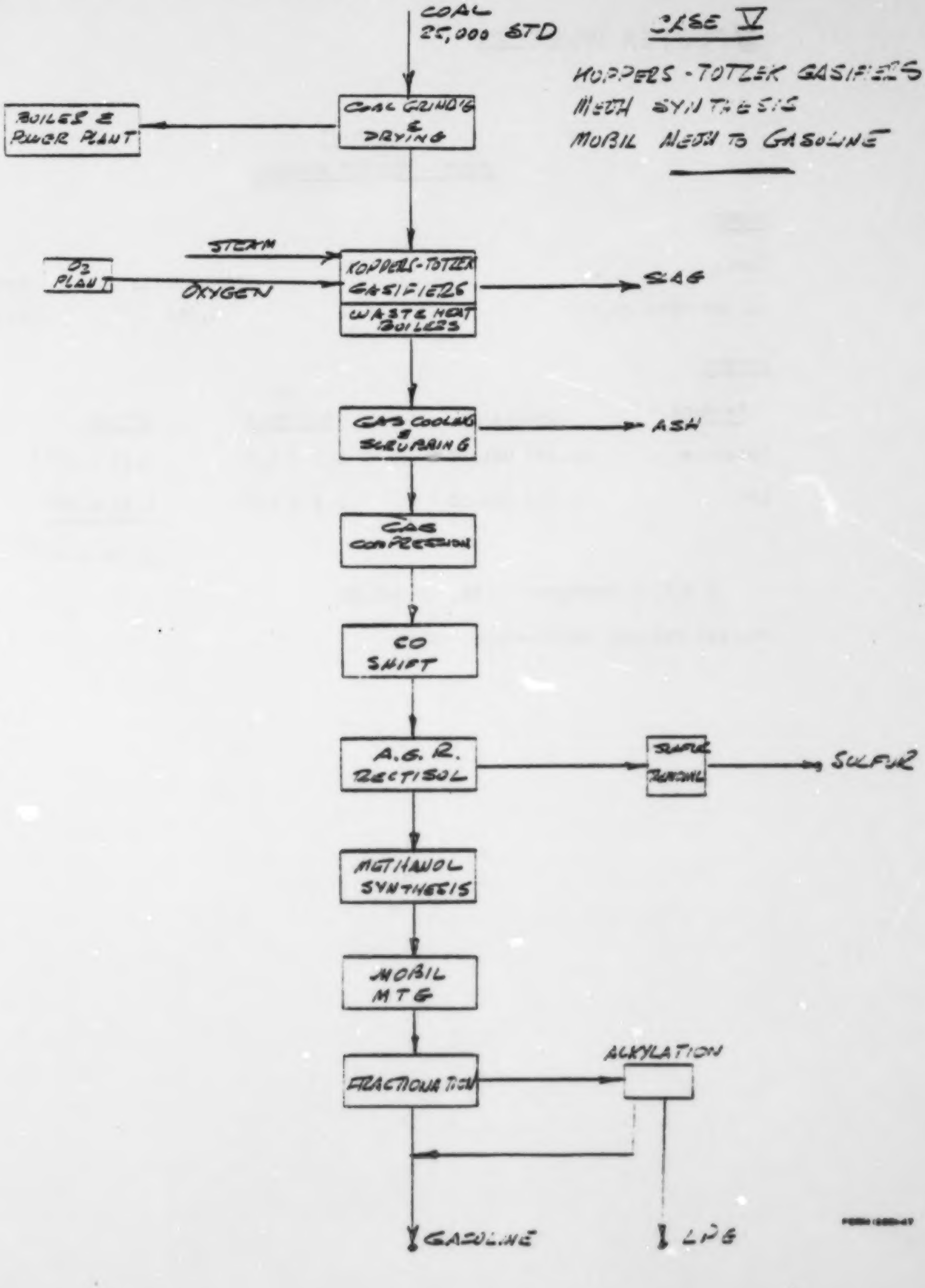
** For an indepth discussion of the effect of slurry concentration and gasifier pressure levels on the economics, the reader is referred to the EPRI Report AF-1161, Sept. 1979, Production of Methanol from Lignite.

The Texaco with the F-T synthesis will provide product yields similar to the ones found for the Winkler - F-T case (Case IV), and it was considered not justifiable for the level of effort allowed for this report to develop these data.

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N. J.
BY G.G. DATE 10/20/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF
CHKD. BY DATE CASE V - K-T TO MTG. JOB NO. 3-28-2011

CASE V

KOPPERS-TOTZEK GASIFIERS
METH SYNTHESIS
MOBIL METH TO GASOLINE



CASE VI

INPUT - OUTPUT SUMMARY

INPUT

Coal:

	HHV,BTU/LB	BTU/SD
25,000 TPSD (A.R.)	6,783	3.392×10^{11}

OUTPUT

Product	Quantity	HHV BTU/Unit	BTU/SD	%
Gasoline	22,357 Bbl/SD	5.1×10^6	1.14×10^{11}	92.95
LPG	2,275 Bbl/SD	3.8×10^6	8.65×10^9	7.05
			1.226×10^{11}	100.00

F.O.E. 6 MMBTU/Bbl : 20,433 Bbl/SD

Overall Thermal Efficiency: 36.2%

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE VII

BRITISH GAS - LURGI (BGC/L) GASIFIER TO GASOLINE VIA METHANOL (MTG)

Process Description*

The lignite is crushed and sized to approximately 1/4" x 2". A portion of the coal fines, -1/4", which are assumed as 25% of the total coal fed to the facility, is used for captive power and process steam generation and the remainder for export power generation.

The sized coal is fed to the gasifiers via lockhoppers. The BGC/L gasifiers are similar in construction to the Lurgi dry bottom type, the main difference being the lack of the ash discharging rotary grate. In the slagging type, the gasifier bottom is fitted with a conical refractory hearth, which has a bottom slag tap hole, which discharges the molten slag into a quench chamber which is followed by a slag lockhopper.

Steam and oxygen are supplied by tuyeres located in the lower section of the gasifier.

The operating fundamental differences between the dry bottoms and slagging gasifiers are the lesser requirements for steam and oxygen in the slagger, the production of a raw gas having a high CO/H_2 ratio and a higher output per unit of a cross sectional area.

The smaller steam requirements result in a lower output of aqueous liquors with consequent decreased waste water treatment requirements.

The thermal efficiency is also greater in the slagging gasifiers.

---

* Refer to attached flowsheet.

The higher steam requirements for the dry bottom type stem from the necessity of tempering the reaction zone temperature to values below the ash softening point.

Other characteristics worth mentioning are the lower CO_2 and methane concentrations in the raw gas, and the reported possibility of the coal fines utilization by injecting them through the tuyeres into the reaction zone.

The BGC has operated a 6' diameter gasifier with reported satisfactory results, and it seems they are now prepared to grant licenses with full commercial guarantees for sizes up to 8' diameter.

The raw gas leaving the gasifier is processed in a similar sequence of operation to the ones described in Case I.

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BY A G DATE 10/8/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. ____ OF ____

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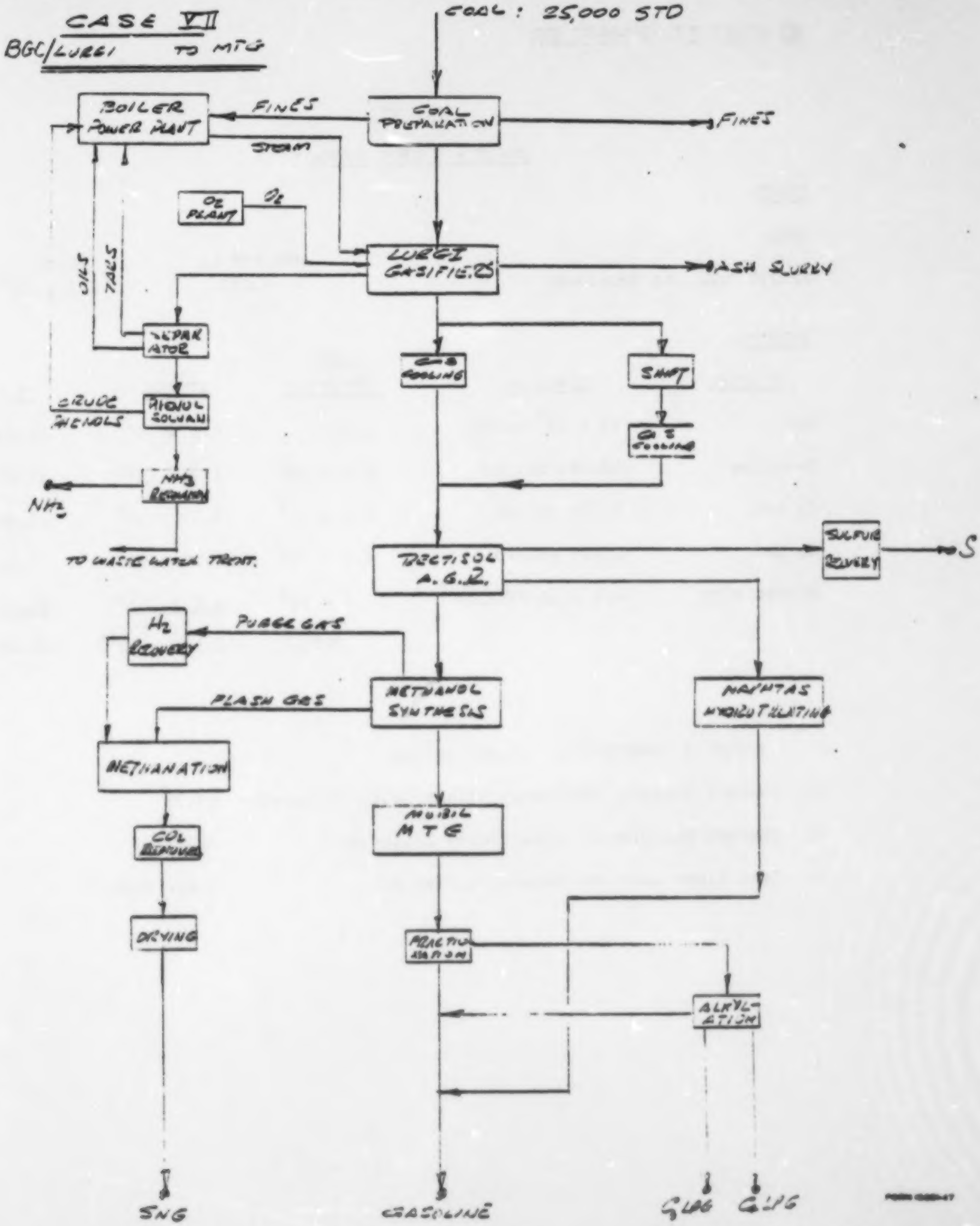
CASE VII

BGC/LURGI TO MTG



Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE VII

INPUT - OUTPUT SUMMARY

INPUT:

Coal:

	HHV, BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

OUTPUTS:

Product	Quantity	HHV BTU/Unit	BTU/SD	%
SNG	52×10^{6} SCFSD	1,000	5.2×10^{10}	24.25
Gasoline	19,593 Bbl/SD	5.1×10^{6}	1.0×10^{11}	46.63
C_3 LPG	1,378 Bbl/SD	3.8×10^{6}	5.23×10^{9}	2.45
C_4 LPG	1,950 Bbl/SD	4.2×10^{6}	8.19×10^{9}	3.82
Electricity	4.9 x 10 KWHR/SD	1×10^{4}	4.9×10^{10}	22.85
		TOTAL:	2.144×10^{11}	100.00

F.O.E. @ 6MMBTU/Bbl: 27,567 Bbl/SD

A. Overall Thermal Efficiency (Electricity Included): 63.2%

B. Thermal Efficiency (Electricity Excluded) : 48.8%

C. Coal Fines Left For Export In Case B: 3,606 TPSD

CASE VIIA

BGC/L GASIFIER TO GASOLINE VIA METHANOL (MTG) WITH SNG REFORMING

Process Description

The flowsheet for this case is similar to Case IA*, with the exception of the gasifier type, which in the present case is the BGC/L slagging bottom. The main differences that should be pointed out are the large shift requirements for the slagging gasifier, and the smaller aqueous liquors and CO_2 produced in the present case.

* Refer to attached flowsheet.

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE VIIA

INPUT - OUTPUT SUMMARY

Input

Coal	HHV, BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	3.392×10^{11}

Output

Product	Quantity	HHV BTU/Unit	BTU/SD	%
Gasoline	27,034 Bbl/SD	5.1×10^6	1.38×10^{11}	69.79
C_3 LPG	1,378 Bbl/SD	3.8×10^6	5.23×10^9	2.64
C_4 LPG	1,950 Bbl/SD	4.2×10^6	8.19×10^9	4.15
Electricity	4.63×10^6 KwHR/SD	1×10^4	4.63×10^{10}	23.42
			1.977×10^{11}	100.00

F.O.E. @ 6 MMBTU/Bbl: 26,170 Bbl/SD

A. Overall Thermal Efficiency (Electricity Included): 58.3%

B. Thermal Efficiency (Electricity Excluded): 44.6%

C. Coal Fines Left for Export in Case B: 3,415 TPSD

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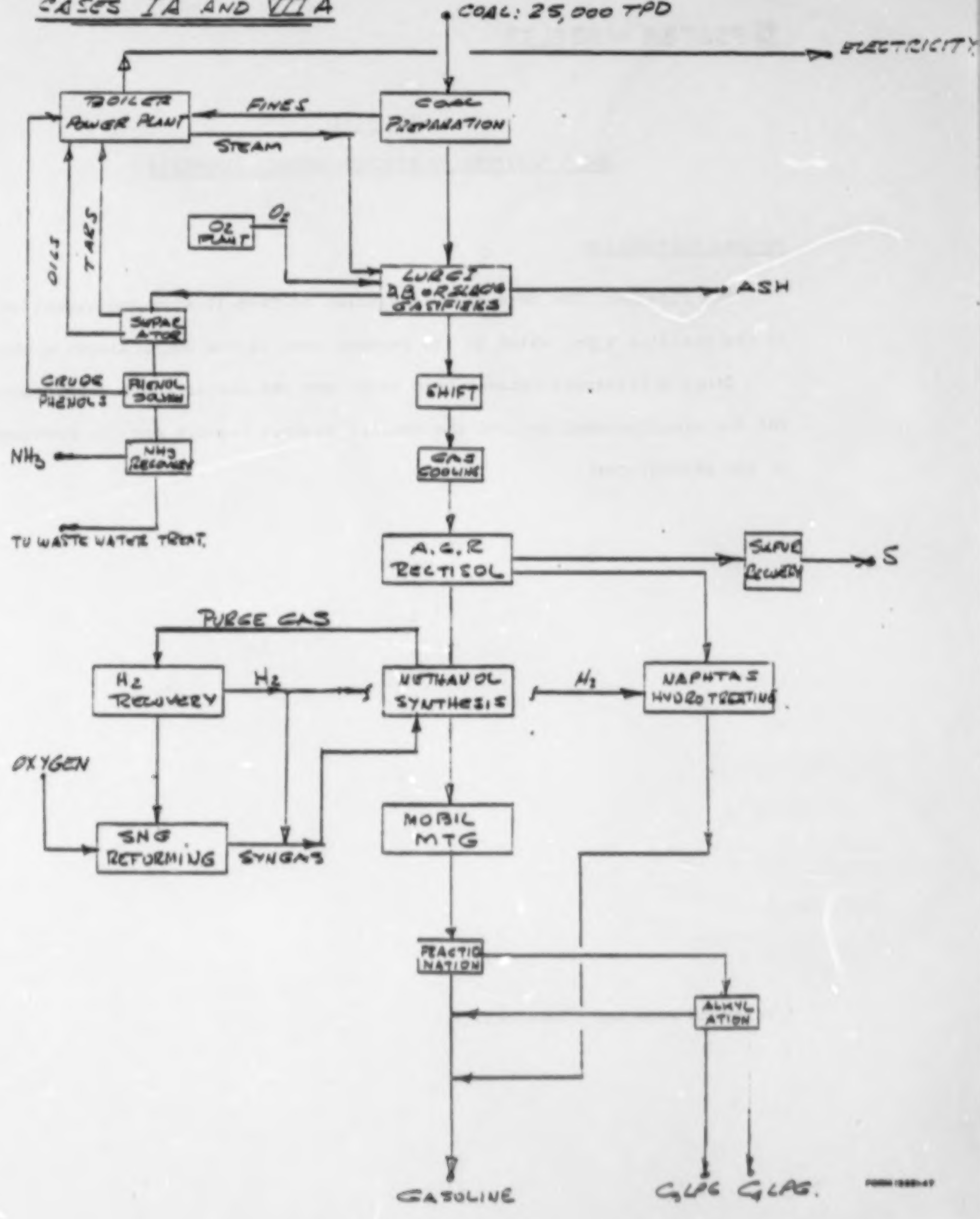
BY A.G DATE 10/27/81 SUBJECT SHEET NO. 1 OF 2

CHKD. BY DATE JOB NO. 3-29-8011

CASES IA AND VIIA



Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE VIII

BGC/L GASIFIER TO FISCHER-TROPSCH SYNTHESIS

Process Description

The flowsheet for this case is similar to Case II* with the exception of the gasifier type, which in the present case is the BGC/L slagging bottom.

Other differences between both cases are the larger shift requirements for the slagging gasifier and the smaller aqueous liquors and CO_2 Produced in the present case.

---

* Refer to attached flowsheet.

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

CASE VIII

INPUT-OUTPUT SUMMARY

Input

Coal:	HHV, BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	$3{,}392 \times 10^{11}$

OUTPUT

Product	Quantity	HHV BTU/Unit	BTU/SD	%
SNG	79.3×10^6 SCFD	1,000	7.93×10^{10}	43.19
Gasoline	11,679 Bbl/SD	5.1×10^6	5.95×10^{10}	32.42
Diesel Fuel	1,984 Bbl/SD	5.3×10^6	1.05×10^{10}	5.72
C_3 LPG	925 Bbl/SD	3.8×10^6	3.61×10^9	1.96
C_4 LPG	125 Bbl/SD	4.2×10^6	5.25×10^8	0.28
Alcohols	438,600 LB/D	13,630	5.98×10^9	3.25
Electricity	2.42×10^6 KwHR/SD	1×10^4	2.42×10^{10}	13.18
			1.836×10^{11}	100.00

F.O.E. @ 6 MMBTU/Bbl: 26,567 Bbl/SD

A. Overall Thermal Efficiency (Electricity Included): 54.1%

B. Thermal Efficiency (Electricity Excluded); 47%

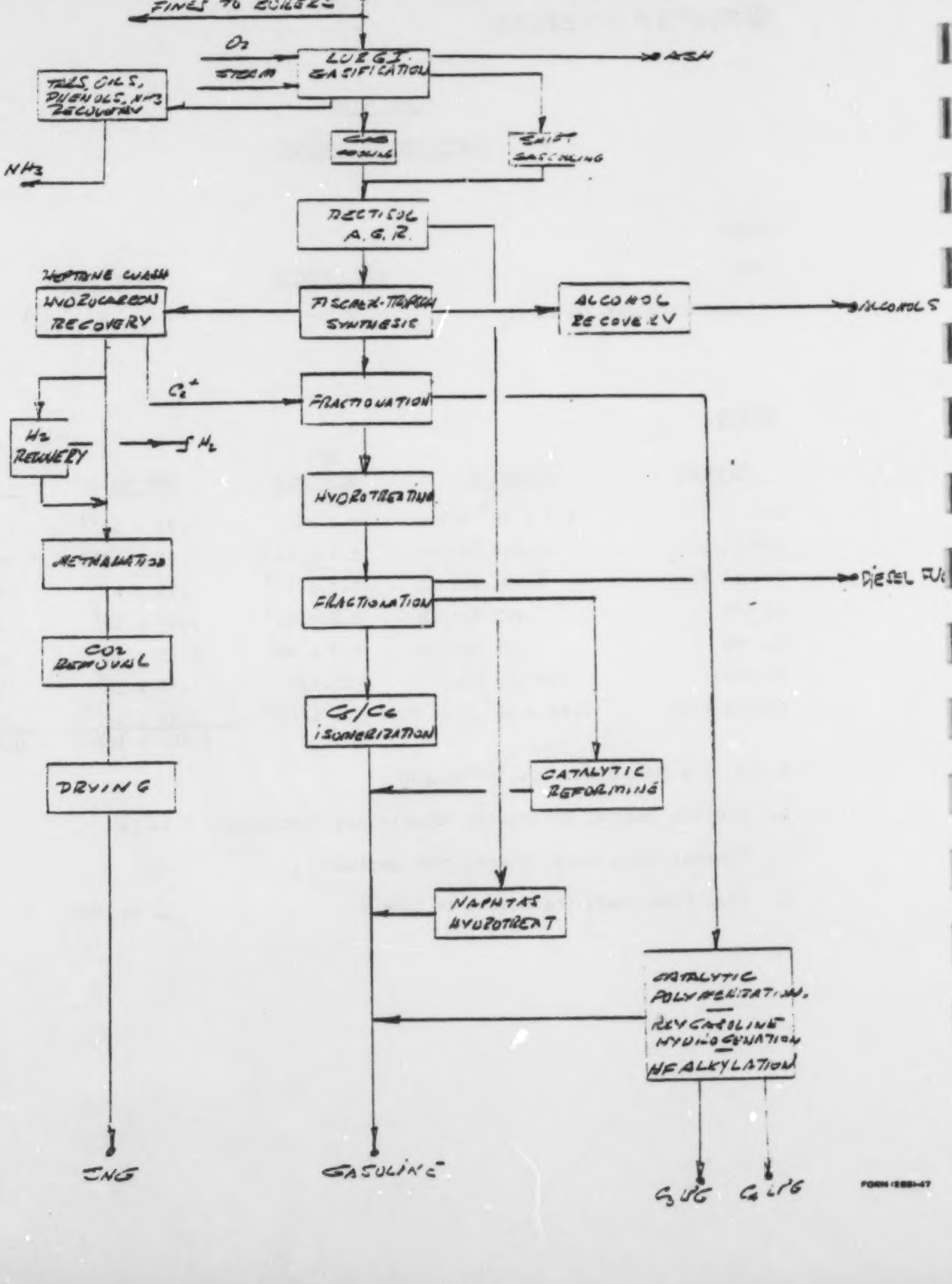
C. Coal Fines Left for Export in Case B: 1,786 TPSD

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N. J.

BY A. G. DATE 10/12/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. ____ OF ____

CHKD. BY ____ DATE CASE II - LURGI (D.B.) → FISCHER-TROPSCH JOB NO. B-28-8011

COAL : 25,000 TPSD



FORM 1288-47

Job No.: B-28-8011
Date: November, 1981

CASE VIIIA

BGC/L GASIFIER TO FISCHER-TROPSCH SYNTHESIS WITH SNG REFORMING

Process Description

The flowsheet for this case is similar to Case IIA,* with the exception of the gasifier type, which in the present case is the BGC/L slagging bottom.

Smaller aqeuous liquor flows, CO_2 and methane are produced in the slagging gasifiers. The ratio of H_2/CO is close to 1/2 and a substantial amount of shifting is required to meet the ratios of about 2 or better for the syngas fed to the F-T synthesis.

* Refer to attached flowsheet.

CASE VIIIA

INPUT - OUTPUT SUMMARY

Input

Coal:	HHV, BTU/LB	BTU/SD
25,000 TPSD (As Received)	6,783	$3,392 \times 10^{11}$

Output

Product	Quantity	HHV BTU/Unit	BTU/SD	%
Gasoline	20,333 Bbl/SD	5.1×10^6	1.37×10^{11}	65.50
Diesel Fuel	3,454 Bbl/SD	5.3×10^6	1.83×10^{10}	11.56
C_3 LPG	1,657 Bbl/SD	3.8×10^6	6.29×10^9	3.97
C_4 LPG	218 Bbl/SD	4.2×10^6	9.16×10^8	0.58
Alcohols	763,627 LB/SD	16,630	1.04×10^{10}	6.57
Electricity	1.87×10^6 KwHR/SD	1×10^4	1.87×10^{10}	11.82
				100.00

F.O.E. @ 6 MMBTU/Bbl: 23,267 Bbl/SD

A. Overall Thermal Efficiency (Electricity Included): 46.7%

B. Thermal Efficiency (Electricity Excluded); 41.2 %

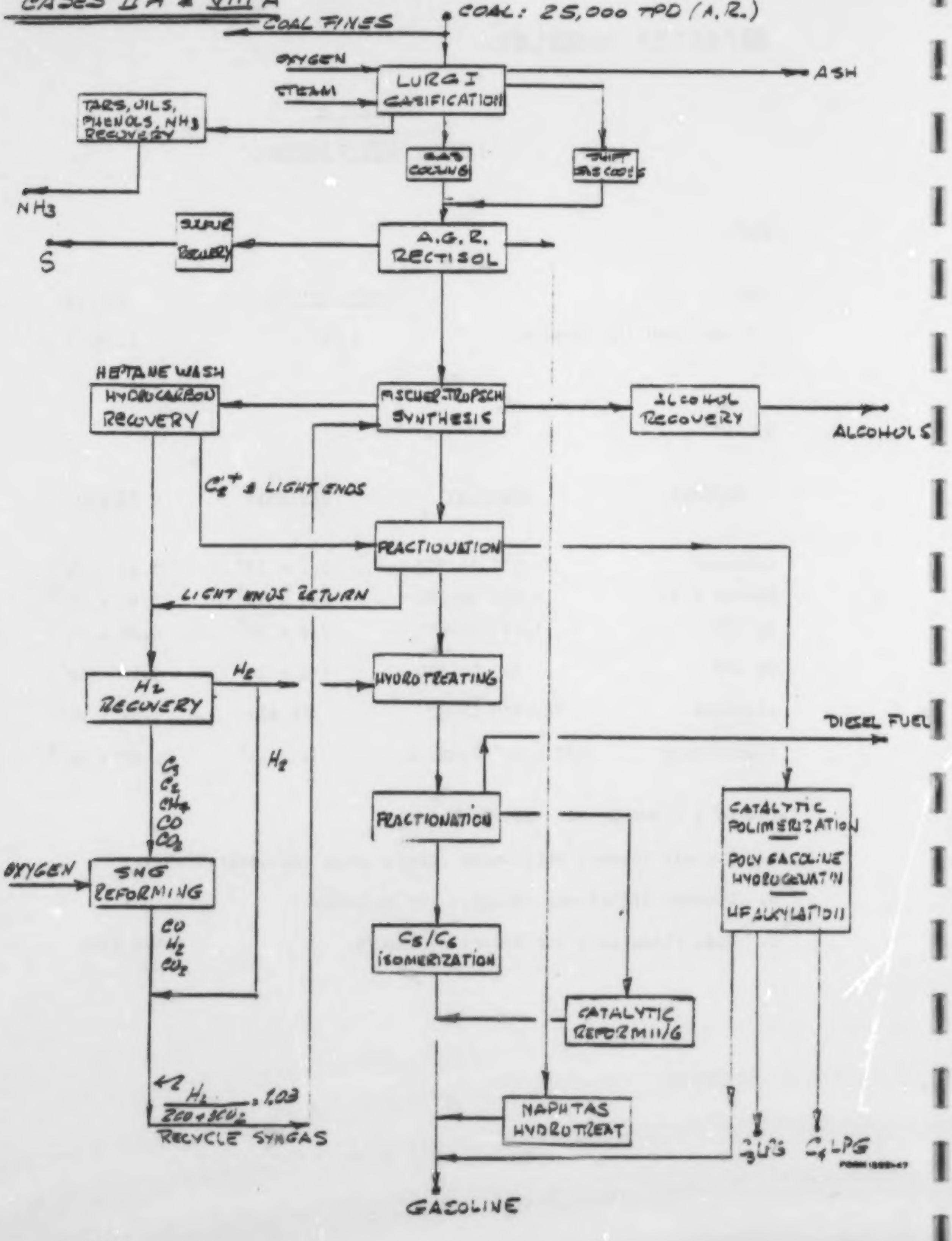
C. Coal Fines Left for Export in Case B: 1,379 TPSD

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N.

BY A.G. DATE 12/21/81 SUBJECT SHEET NO. 2 OF 2

CHKD. BY DATE JOB NO B-28-8011

CASES IIA & VIII A



EXPECTED PRODUCTS QUALITY

EXPECTED PRODUCTS QUALITY

GASOLINE FROM METHANOL BY THE MRDC PROCESS

COMPARISON OF PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS WITH ESTIMATED METHANOL-TO-GASOLINE GASOLINE PROPERTIES-

	Estimated MTG Unleaded Gasoline Properties	Specification
Gravity, °API	61.4	
Octane Numbers		
Research	93	
Motor	83	82 min.
(Research + Motor)/2	88	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation, °F		
IBP	85.5	
10%	110.5	158 max.
30%	146.1	
50%	200.0	170/250 min./max
70%	262.8	
90%	336.9	374 max.
EP	388.6	437 max.
V/L Ratio (=20), °F	@ 129.5	@ 140 max.
Sulfur, wt. %	Nil	0.10 max.
Composition, vol. %		
Paraffins	51.3	
Olefins	11.3	20 max. (target)
Naphthenes	8.7	
Aromatics	28.7	
Durene Content, wt. %	4.6	5 max. (target)
Molecular Weight	94.3	

SOURCE: D.O.E. FE-2447-9, Jan. 1978

PRODUCTS QUALITY

SNG: BY LURGI (D.B.) TO GASOLINE VIA METHANOL

Composition

Hydrogen	1.7
Methane	95.5
Ethane	0.2
Propane	0.1
Butane	0.1
Carbon dioxide	0.5
Inerts (N_2 + Ar)	1.9
	100.0

Other

Heat of combustion (HHV)	980 Btu/SCF
Carbon monoxide (0.1% max.)	0.02%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

Index	Calculated	Preferable	Objectionable
Lifting, I_l	1.03	under 1.0	above 1.06
Flash-back, I_f	1.02	under 1.15	above 1.2
Yellow-tip, I_y	1.05	above 1.0	under 0.8

The SNG product is of satisfactory quality and is compatible with pure methane. The inerts from the oxygen stream for gasification, however, give rise to a lifting index value between the preferable and objectionable values. From a processing standpoint, there is little that can be done to reduce this value. The selection of a more expensive oxygen plant yielding a 99+% pure product, however, would alleviate this problem. (The SNG product, on the other hand, has a lifting index value better than the preferable value when compared with the high methane natural gas standard cited in the A.G.A. Research Bulletin Number 36. This gas has a 5.0% inerts content.)

SOURCE: D.O.E. FE-2447-9, Jan. 1978

PRODUCTS QUALITY

C_3LPG & C_4LPG

COMPARISON OF PRINCIPAL LPG SPECIFICATIONS WITH ESTIMATED METHANOL-TO-GASOLINE LPG PROPERTIES-

	Commercial Propane ASTM D1835*	Est. MTG C_3 LPG Properties	Commercial Butane ASTM D1835*	Est. MTG C_4 LPG Properties
Vapor Pressure at 100°F, max. psig (D1267)	208	171	70	52
Volatile Residue				
Butane and Heavier, max. vol. %	2.5	1.9	-	-
Pentane and Heavier, max. vol. %	-	-	2.0	0.7
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

*1976 Annual Book of ASTM Standards

SOURCE: P.O.E. FE-2447-9, Jan. 1978

COMPARATIVE INVESTMENT COSTS

COMPARATIVE INVESTMENT COSTS

An attempt has been made to place the investment costs for the various cases considered in this report under a common baseline.

The emphasis has been on their relative values rather than their absolute values, which carry no claim of accuracy. This would require an unduly and possible not justifiable effort for the purposes of this report.

Data from published references and in-house data has been used throughout.

The costs are reported in 4th quarter 1981 dollars:

Case I : 2,000 MM$

Case IA: 2,100 MM$

The increase over Case I reflects the larger oxygen plant capacity due to the requirements of the Partial-Oxidation of the SNG, the addition of the Partial-Oxidation Plant, the extra boiler capacity, the increase in the Methanol Synthesis Train and the Power Plant for the coal fines.

Case II: 2,100 MM$

The higher complexity of the Fischer-Tropsch Synthesis as compared with the methanol to gasoline synthesis (Case I) explains the cost differential.

Case IIA: 2,300 MM$

The higher investment cost for this case is due to reasons similar to Case IA.

Case III: 1,600 MM$

It has to be noticed that the pressure in the Winkler gasifier was assumed to be 60 psig.

Failure to achieve successful operation at the higher pressure will result in an increase of the number of gasification trains and the compression plant, with an expected additional investment of about 20%.

Case IV : 1,750 MM$

The higher figure for Case IV as compared to Case III is a consequence of the use of F-T synthesis instead of the methanol to gasoline process.

Case V : 1,800 MM$
Case VI : 1,500 MM$

This case assumes the Texaco gasifiers operating at 1,500 psig, and feeding a 43% coal slurry into them. (Reference: EPRI AF-1161, Sept. 1979(

The high pressure level was selected to improve the overall efficiency when combined with a high pressure methanol synthesis plant. This pressure level (1,550 psig) is an extrapolation beyond the present experience of Texaco's pilot plant gasifier at Montebello, California. The present coal gasifiers at Texaco's faciliy is designed for 1,200 psig.

An increase in the investment would result in the lower pressure case. The amount of this increase has not been estimated.

Case VII : 2,250 MM$

In this case, which is similar to Case I with the exception of the use of the BGC/L gasifiers instead of the Lurgi D.B., the higher efficiency of the BGC/L gasifiers generates larger amounts of syngas which require a corresponding larger plant capacity.

This fact explains the cost differences between both cases.

Case VIIA : 2,300 MM$
Case VIII : 2,350 MM$
Case VIIIA: 2,550 MM$

The reasons for the costs [illegible]nentials in these cases are similar to the ones mentioned for Case IA and II.

COMMENTS

COMMENTS

The salient characteristics of the various gasifiers and processes considered in this report have been described in the process descriptions.

It should be noticed that due to the basic assumptions made, all the cases dealing with the Lurgi or British Gas-Lurgi gasifiers show an export of electric power which is due to the excess coal fines. The amount of power generated varies between 100 MW and 200 MW, depending on the particular case considered.

An important consideration in the process selection will be the capability of the local electric grid to absorb this amount of power.

Another significant point is the relatively large production of SNG by the Lrugi and BGC/L gasifiers as compared with the Winkler, Koppers-Totzek and Texaco, which produce mininal, if any, SNG.

The requirements of a pipeline, with its associated costs, for the delivery of the SNG will be another of the factors that will enter in the process selection.

The quality of the gasoline and SNG, as well as diesel fuel and LPG's is expected to be within commercial specifications. See attached "products quality" tables.

The following brief discussion is intended to clarify some apparent discrepancies when comparing thermal efficiencies for the various cases.

The BGC/L slagging gasifier is reported to have a higher thermal efficiency than the Lurgi dry bottoms. When comparing two similar cases: I and VII, the found overall efficiencies are 52.8% for Lurgi DB case and 48.8% for the BGC/L case.

This would contradict the fact mentioned above. This apparent discrepancy is due to the larger amount of CH_4 generated in the Lurgi DB, 92×10^6 SCF/SD as compared to 52×10^6 SCF/SD in the BGC/L.

Although the gasoline product in the BGC/L is larger than the Lurgi's DB: 19,593 Bbl/SD for BGC/L and 15,07 Bbl/SD for Lurgi DB, the larger heating value per unit carbon for methane than for gasoline is responsible for the greater overall thermal efficiency. To illustrate this point, the carbon used up for the CH_4 and for gasoline, per pound of carbon in the feed is:

Case I

	Fraction of Carbon in Products	HHV Btu/Lb Carbon	Btu/Lb Carbon Fed
SNG:	0.217	32,000	6,944
Gasoline:	0.230	23,400	5,382
		Total	12,326

Case VII

	Fraction of Carbon in Products	HHV Btu/Lb Carbon	Btu/Lb Carbon Fed
SNG:	0.214	32,000	3,968
Gasoline:	0.32	23,400	7,488
		Total	11,456

$$\text{Ratio of } \frac{\text{Case I}}{\text{Case VII}} = \frac{12{,}326}{11{,}456} = 1.08$$

Ratio of Overall Efficiencies:

$$\frac{\text{Case I}}{\text{Case VII}} = \frac{52.8}{48.8} = 1.082$$

It can now be seen that the influence of the SNG on the overall heat output is predominant.

The maximization of liquids fuels requires the conversion of the SNG into Syngas, which in this study has been assumed to be performed by partial oxidation of the SNG.

Although the overall liquids yield is increased, the overall thermal efficiency decreases quite substantially. This is due to the additional inefficiencies incurred, and mainly, to the incidence in the overall heat output of the high heating values for methane, as explained above.

The Case III, which utilizes a Winkler gasifier under 60 psia, when in combination with a methanol to gasoline synthesis plant produces one of the highest yields of gasoline (24,067 Bbl/SD), only surpassed by Case VIIA (27,034 Bbl/SD).

This relatively high yield is a consequence of the gasifier higher working pressure, which by decreasing the compression power requirements renders a satisfactory performance.

It has to be noticed that the assumed operating pressure has not yet been used in any commercial operation, which work at atmospheric pressure. Failure to achieve the higher pressure will result in a larger number of gasifiers and an increased gas compression plant, which will result in a higher capital expenditure and lesser gasoline yield.

Case V, which uses a Kopper-Totzek gasifier, also shows a relatively high gasoline yield (23,255 Bbl/SD). A characteristic of this type of gasification is the large compression plant requirement coupled with relatively high oxygen requirements.

The carbon utilization for lignites has been reported at 98%, which is excellent, specially if compared with about 85% for the Winkler.

A feature of the K-T is the generation of a raw gas containing a small percentage of CO_2 and a CO/H_2 ratio of about 2. This requires a larger shifting plant than the Lurgi dry bottoms.

Due to the high temperature of the gas leaving the gasifier, the generation of steam from its heat content is quite substantial.

Nevertheless, this is counter balanced by the relative high power requirements of the K-T gasification.

The Texaco gasifier, which also shows a good yield of gasoline (22,367 Bbl/D), has been assumed to operate at 1500 psig which is higher than the maximum 1200 psig tested at the Texaco Pilot Plant in Montebello, California.

There is as yet no commercial experience with this type of gasifier for coal, although there is extensive commercial experience for liquid fuels partial oxidation. The cost slurry concentration assumed here is 43%, and this is also a critical parameter for the plant cost and power input.

If greater slurry concentrations would be considered, the overall thermal efficiency would show a gain. It has to be noticed that operations at lower pressures would increase the number of gasification trains, and the power input with the concomitant increase in capital cost.

British Gas-Lurgi slagging bottom gasifiers, as considered in Case VII and subsequents, show good yields of gasoline, specially in Case VIIIA (27,035 Bbl/SD). They also produce substantial amounts of diesel fuel and alcohols, which is common to all F-T synthesis. When the investment cost per daily barrel are compared between similar cases for the Lurgi dry bottom and the BGC/L, the last one shows a lower cost. This is a reflection of the higher thermal efficiency for the BGC/L gasifiers.

The suitability of the various gasifiers considered in this report for lignite will now be briefly discussed.

The Lurgi dry bottoms gasifier is said to be suitable for most types and ranks of coal.

Ash properties and coal reactivity are the main factors that determine the ratio of steam to oxygen fed into the gasifier.

Moisture content of the coal is not particularly critical, provided it does not reduce the temperature of the leaving gas below the dew point.

German lignites have been successfully gasified, and various plants have been proposed for coal to SNG in the U.S., one of them dealing with a hard lignite from the North Dakota area (American Natural Gas, Beulah-Hazen, N.D.).

It would thus seem that the Saskatchewan lignites could be handled by the Lurgi d.b., although here as in any other case, tests should be conducted with the actual coal to determine the suitability of the particular gasifier under consideration.

The British Gas-Lurgi slagging bottoms, has been tested, not on commercial scale, on various types of coals, but it is not known if a lignite has been tested, although it would appear at the surface that no unsurmountable difficulties should be encountered for lignites.

The Winkler gasifiers can handle a variety of coals ranging from brown coal to moderately caking coals. It cannot handle coals that have a very low ash melting point and very low reactivity.

These gasifiers have handled German and other procedence lignites, and would appear certainly suitable for the Saskatchewan lignites.

Currently they are still left some operating plants that generate synthesis gas for ammonia production (Kuthaya, Turkey; Madras, India; Gorsade, Yugoslavia).

Koppers-Totzek gasifiers can handle virtually any kind of coal. They have been in commercial operation since the early fifties with about 50 gasifiers over 14 countries. Most of these plants product a synthesis gas for ammonia.

It is reported that these gasifiers have operated successfully with high ash lignites (Corina, Spain; Kuthaya, Turkey).

It would seem that they could handle satisfactorily the proposed lignite.

The Texaco Development Corporation reported to C.F. Braun Co. (see page 4-4 of Part II of the EPRI Report No. AF-1161) that they have run a 55% of pretreated lignite slurry at 300 psig.

The lignite used was from the Knife River Coal Co. of the South Beulah, N.D., deposit.

This fact would indicate that, in principle, the feasibility of gasifying a lignite in the Texaco gasifiers is established.

Further testing at higher pressures contemplated, 1200 psig or higher should be performed with the actual coal under consideration before any firm statement be made on the suitability aspect.

300

STATUS

Lurgi, Dry Bottom: Commercial, largest unit: 13 ft. diameter, gasification rates about 150 lb. coal per hour per ft^2.

Winkler: Commercial at atmospheric pressure operation, 45 psig pressure is in the development stage.

Largest commercial unit: 18 ft. diameter, capable of gasifying about 900 tons per day of coal.

Koppers-Totzek: Commercial, largest unit: 4 burner heads capable of gasifying about 800 TPD of coal, producing about 43 MMSCFD of gas.

Texaco: Under development, largest unit is a pilot plant at Oberhausen, West Germany handling 150 TPD of coal.

British Gas-Lurgi Slagging: Under development, largest unit is 6 ft. diameter with gasification rates of about 600 to 800 lb. coal per hour per ft^2.

Fischer-Tropsch Synthesis: Commercial, largest units at Sasol.

Methanol-to-Gasoline Synthesis: Under development, MRDC operated pilot plant.

Job. No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

GASIFIERS TEMPERATURE RANGE

	Temperature Range
Lurgi Dry Bottom	Below ash softening temperature, about 1900°F, at the bottom of the gasifier to about 570°F to 1200°F at the top of the gasifier.
Winkler	1500°F to 1900°F
Koppers-Totzek	3000°F to 3500°F
Texaco	2000°F to 2800°F
BGC/L Slagging	700°F to 1200°F at the top 2500°F at the bottom

TABULATION OF SIGNIFICANT RESULTS

FOSTER WHEELER SYNFUELS CORPORATION

Job No. B-28-8011
Date: November, 1981

TABLE I

TABULATION OF SIGNIFICANT RESULTS

Prescreening Study

Case	I	IA	II	IIA	III	IV	V	VI	VII	VIIA	VIII	VIIIA
Gasifier	Lurgi Dry Bottom	Lurgi Dry Bottom	Lurgi Dry Bottom	Lurgi Dry Bottom	Winkler	Winkler	Koppers Totzek	Texaco	British Gas Lurgi	British Gas Lurgi	British Gas Lurgi	British Gas Lurgi
Liquefaction	MeOH MTG	MeOH/MTG SNG. Ref.	F-T	F-T SNG. Ref	MeOH MTG	F-T	MeOH MTG	MeOH MTG	MeOH MTG	MeOH/MTG SNG. Ref.	F-T	F-T SNG. Ref.
Gasifier Pressure, psig	400	400	400	400	45	45	6	1,550	400	400	400	400
Overall Thermal Efficiency Electricity Included, %	66.6	53.8	57.4	41.0	N.A.	N.A.	N.A.	N.A.	63.2	58.3	54.1	46.7
Overall Thermal Efficiency Electricity Excluded, %	52.8	41.0	48.7	33.8	39.0	38.0	37.0	36.2	48.8	44.6	47.0	41.2
Gasoline Yield. Bbl/SD	15,070	25,190	8,915	16,712	24,067	10,786	23,255	22,357	19,593	27,034	11,679	20,333
Export Power, KWHR/SD	4.71×10^6	4.36×10^6	2.93×10^6	2.4×10^6	—	—	—	—	4.9×10^6	4.63×10^6	2.42×10^6	1.87×10^6
Product Slate in % of Overall Efficiency												
SNG, %	40.76	—	53.44	—	—	40.56	—	—	24.25	—	43.19	—
Gasoline, %	33.96	70.45	23.38	61.39	92.78	42.74	92.78	92.95	46.63	69.79	32.42	65.50
Diesel Fuel, %	—	—	4.13	10.88	—	8.47	—	—	—	—	5.72	11.56
C_3 LPG, %	1.77	2.20	1.44	3.73	7.22	2.93	7.22	7.05	2.45	2.64	1.96	3.97
C_4 LPG, %	2.78	3.45	0.21	0.55	—	0.45	—	—	3.82	4.15	0.28	0.58
Alcohols, %	—	—	2.35	6.16	—	4.83	—	—	—	—	3.25	6.57
Electricity, %	20.82	23.90	15.05	17.29	—	—	—	—	22.85	23.42	13.18	11.82
Fuel Oil Equivalent, Bbl/SD	29,853	23,137	27,550	19,128	22,033	21,450	20,917	20,433	27,567	26,170	26,567	23,267
Syngas to Synthesis, LBM/SD	53,298	84,354	53,716	92,691	69,390	63,393	65,880	63,350	64,323	88,492	64,445	107,541
Investment Cost, MM$	2,000	2,100	2,100	2,300	1,600	1,750	1,800	1,500	2,250	2,300	2,350	2,550
Investment Cost Per Daily Bbl of Gasoline, $	132,710	99,103	253,560	137,630	66,481	162,250	77,403	67,073	114,840	85,078	201,220	125,410
Investment Cost Per Daily Bbl of F.O.E., $	66,995	90,764	76,225	120,240	72,618	81,585	86,054	73,411	81,619	87,887	88,456	109,600
Commercial Status, Year*	1985	1985	1981	1981	1985	1981**	1985***	1985	1985	1985	1981	1981

* Estimate
**Gasifier at Atmospheric Pressure
***Would be 1981 if Fischer-Tropsch Synthesis.

DISCUSSION AND CONCLUSIONS

DISCUSSION AND CONCLUSIONS

As stated in the "Selection Criteria" part of this report one of the factors that will strongly influence the process selection is the marketability of the products.

It is not known at the present stage of the project which products are marketable and what prices they would command.

For the pruposes of this discussion, certain hypothetical assumptions will be made in order to show the general procedure for selection.

The first assumption will be that all products are saleable and that the price will be based only on their high heating value, irrespective of the nature of the product.

Under this assumption, the lowest investment cost per daily F.O.E. barrel will indicate the most promising case.

From the inspection of the tabulated values for these parameter in Table I is may be seen that Case I is the most favorable ($66,995) and is closely followed by Case III ($72,618), case VI ($73,411) and Case II ($76,225).

The time frame reference for the commercial status criteria would eliminate Cases I, III and VI if instant execution of the project is considered.

As a second assumption it will be considered that only liquid products are marketable.

This assumption eliminates Cases I, II, IV, VII and VIII, which generate sizable quantities of SNG.

If further, the sales price for the liquid products is assumed uniform and only based on their high heating values, then the following cases will have the lowest investment per F.O.E. daily barrel:

Case III :	Winkler - MeOH - MTG	($72,618)
Case VI :	Texaco - MeOH - MTG	($73,411)
Case V :	Koppers-Totzek - MeOH - MTG	($86,054)
Case VIIA:	BGC/L - MeOH - MTG	($87,887)

The comments made in another part of this report indicate that the calculated F.O.E. daily barrel investment cost for the Winkler and Texaco gasifiers may be optimistic.

The commercial status would favor Cases V and VIIA, due to the experience with K-T and BGC/L gasifiers.

It may now be realized that the final selection of the most promising case will have to await till the necessary background information is made available.

REFERENCES

1. R. B. Sharman and J. E. Scott, "The British/Gas Lurgi Gasifier and Its Relevance for Power Generation," to be presented at the EPRI Conference on Synthetic Fuels, San Francisco, undated.

2. David Gray, Milton Lytton, Martin Newworth, Glen Tomlinson, "The Impact of Developing Technology on Indirect Liquefaction, D.O.E./ER-10280-TI, November, 1980.

3. Max Schreiner, "Research Guidance Studies to Assess Gasoline from Coal by Methanol-to-Gasoline and Sasol Type Fischer-Tropsch Technologies," D.O.E. No. FE-2447-13, August, 1978.

4. "Production of Methanol from Lignite," EPRI AF-1161, September, 1979.

5. "Lignite to Methanol, An Engineering Evaluation of Winkler Gasification and ICI Methanol Synthesis Route," EPRI, AP-1592, October, 1980.

6. "Coal to Methanol, An Engineering Evaluation of Texaco Gasification and ICI Methnaol Synthesis Route", EPRI AP-1962, August, 1981.

Volume

4

STUDY RELATING TO THE TECHNICAL AND ECONOMIC FEASIBILITY OF LIGNITE LIQUEFACTION IN SASKATCHEWAN

Volume IV

REVIEW OF COAL LIQUEFACTION TECHNOLOGIES AND CONCEPTS CONSIDERED FOR COMMERCIALIZATION AROUND 1995

Final Report
by

TECHNICAL DEVELOPMENT DEPARTMENT, SNC Inc.

for

SASKATCHEWAN OIL AND GAS CORPORATION

and

CANMET, ENERGY, MINES AND RESOURCES, CANADA

October, 1982
Montreal

REQ. NO. 2-9203

STUDY RELATING TO THE TECHNICAL AND ECONOMIC FEASIBILITY OF LIGNITE LIQUEFACTION IN SASKATCHEWAN

CONTRACT REPORT NO. OSQ82-00269

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J.A. Mikhlin, Eng, Ph.D.

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CHAPTER - 1

INTRODUCTION

1.0 INTRODUCTION

This study was carried out by the Coal Conversion Group, Technical Development Department of SNC Inc. to evaluate various coal liquefaction processing alternatives for two designated Saskatchewan lignites, i.e., from the Estevan and Willow Bunch deposits. The processes evaluated were subdivided into two categories.

- processes considered for commercialization by 1990, and
- processes considered for commercialization around 1995.

The scope of the work for the advanced processes (commercialization by 1990) included review and evaluation of direct and indirect coal liquefaction and pyrolysis processes listed in Table 1-1. The evaluation was made in two stages. Stage I comprises of technical assessment, prescreening and ranking of the processes for both Estevan and Willow Bunch lignites. Stage II covers techno-economic assessment of processes selected in Stage I but for Willow Bunch lignite only. The ultimate objective is to determine promising process(es) for Saskatchewan lignites.

The scope of the work in the second category (commercialization around 1995) is to ascertain state-of-the-art technology and review update of the processes and concepts listed in Table 1-2.

In addition, an independant experimental program was carried out concurrently with this study to determine comparative reactivity of Estevan and Willow Bunch (Coronach) lignite. These tests were made by the Chemical Technology Division of B.C. Research and the results are presented in a separate report, "Liquefaction and Pyrolysis Tests on Samples of Estevan and Coronach Lignites", Project Report No. 2-02-691, March 1982. A summary of the results is presented in volume I.

Table 1-1 Processes Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization by 1990

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
H-Coal	Hydrocarbon Research Inc.	U.S.A.
RAG/Veba	Ruhrkohle AG and Veba Oel AG	F.R.G.
Rheinbraun	Rheinische Braunkohlenwerke AG	F.R.G.
Exxon Donor Solvent (EDS)	Exxon Research and Engineering Company	U.S.A.
Liquid Solvent Extraction (LSE)	National Coal Board, Coal Research Establishment	U.K.
	COAL PYROLYSIS	
Lurgi-Ruhrgas	Lurgi Mineralöltechnik GmbH and Ruhrgas AG	F.R.G.
	INDIRECT COAL LIQUEFACTION	
SASOL-Type Fischer-Tropsch	SASOL Ltd.	South Africa
Mobil Methanol-to-Gasoline (M-MTG)	Mobil Research and Development Corporation	U.S.A.
	Gasifiers for Production of Synthesis Gas	
Lurgi (dry bottom)	Lurgi Mineralöltechnik AG	F.R.G.
British Gas/Lurgi (slagging)	British Gas Corporation and Lurgi Mineralöltechnik AG	U.K. F.R.G.
Koppers-Totzek	Krupp-Koppers GmbH	F.R.G.
Texaco	Texaco Research and Development Corporation	U.S.A.
Winkler (high temperature)	Rheinische Braunkohlenwerke AG	F.R.G.
Westinghouse	Westinghouse Electric Corporation	U.S.A.

Table 1-2 Processes and Concepts Selected by the Saskatchewan Coal Liquefaction Committee and Considered for Commercialization around 1995

Process	Process Developer (Licenser)	Country
	DIRECT COAL LIQUEFACTION	
Zinc Halide	Consolidation Coal Co.	U.S.A.
Dow Coal Liquefaction	Dow Chemicals Co.	U.S.A.
Consol Synthetic Fuel (CFF)	Consolidation Coal Co.	U.S.A.
Solvent Refined Coal (SRC-1)	International Coal Refining Co.	U.S.A.
Solvent Refined Lignite (SRL)	University of North Dakota	U.S.A.
CO-Steam	Grand Forks Energy Technology Center	U.S.A.
Two-Step Liquefaction	Cities Service Co. C-E Lummus Co.	U.S.A.
	COAL PYROLYSIS	
Occidental Pyrolysis	Occidental Research Corporation	U.S.A.
	INDIRECT COAL LIQUEFACTION	
	Gasifiers for Production of Synthesis Gas	
Shell Gasifier	Shell International Petroleum	The Netherlands
U-Gas	Institute of Gas Technology	U.S.A.
	CO-PROCESSING	
Coil	Hydrocarbon Research Inc.	U.S.A.
Solvolysis	Mitsubishi Heavy Industries Ltd.	Japan
Veba Coal/Oel Co-processing	Veba Oel AG	F.R.G.

1.1 STUDY REPORT ORGANIZATION

The study report comprises five volumes:

Volume I - Summary and Process Comparison

Volume II - Assessment of Advanced Direct Coal Liquefaction and Pyrolysis Processes Considered for Commercialization by 1990

Volume III - Assessment of Advanced Indirect Coal Liquefaction Processess Considered for Commercialization by 1990

Volume IV - Review of Coal Liquefaction Technologies and Concepts Considered for Commercialization around 1995

Volume V - Process Developer's Data and Information

This volume IV presents a review up-date of coal liquefaction processes and concepts that have not yet been demonstrated (except SRC-I process) on a large pilot-plant scale and, therefore, might be considered for commercialization not earlier than around 1995. The gasifier configurations listed in Table 1-2 were evaluated by Foster Wheeler, U.S.A., and their report to SNC Inc. is attached separately in Annex A of this volume. The FW report also includes Westinghouse gasification process, which was later on considered for commercialization around 1990, in view of the accelerated development program initiated by the process at the end of 1981 - beginning 1982 and a decision to install a 1200-t/d commercial unit at SASOL-II complex by 1985.

1.1 STUDY REPORT ORGANIZATION (cont'd)

The main objectives of the review up-date are as follows:

- to determine the current state-of-the-art for each of the processes considered
- to determine the prospects for further developments (including financial support expected for future work)
- to evaluate the applicability of process concept and/or its particular characteristic features with respect to liquefaction of Saskatchewan lignites.

The processes selected for evaluation have significant differences in concept, approach, and state of development. Evaluation of each of the processes includes history of development, types of coal tested, reactor yields and range of operating parameters established through various experimental stages such as bench-scale program, PDU operation (where applicable, etc.). Also, review of a conceptual design of commercial plant was carried out including overall and liquid product efficiencies (for the processes where such a design has already been done and information is available to SNC). Current status and prospects including financial support expected for future developments of each process is presented. Finally, the applicability of process concepts and/or their characteristic features for future work on liquefaction of Saskatchewan lignites is discussed.

1.0 STUDY REPORT ORGANIZATION (cont'd)

In order to develop the review up-date, the emphasis was given on direct contacts with process developers and visits to their experimental facilities. In addition, reports issued by process developers, U.S. Department of Energy, Electrical Power Research Institute and other organizations concerned, as well as presentations made on the subject at recent conferences, meetings and symposia have been used as basic sources for the study.

CHAPTER - 2

REVIEW OF RECENT DEVELOPMENTS IN SELECTED DIRECT COAL LIQUEFACTION PROCESSES

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2.0 REVIEW OF RECENT DEVELOPMENTS IN SELECTED DIRECT COAL LIQUEFACTION PROCESSES

In this chapter, review up-date on the following seven "direct" coal liquefaction processes selected by the Study Committee is presented:

- Dow Coal Liquefaction
- CO-Steam
- Consol Synthetic Fuel (CSF)
- Two-Stage Liquefaction (TSL)
- Solvent Refined Coal (SRC)
- Zinc-Halide (Z-H)
- Solvent Refined Lignite (SRL)

The first two processes may be referred to as one-step direct liquefaction since coal is hydrogenated (liquefied) in a one-stage reactor. The remaining five processes may be considered as two-step direct liquefaction. In these processes, the first step involves conversion of coal into liquid (or extract in case of SRC and SRL) which is then hydrogenated to yield synthetic crude oil through a second-stage hydrogenation. The liquid from the first-stage is separated from ash and unconverted coal before the second-stage operation. In the SRC and SRL processes, first-stage hydrogenation followed by solids separation yields ash and sulfur-free coal extract, so-called "solvent refined coal" or "solvent refined lignite" product. The SRC product which is solid at room temperature can be hydrocracked to yield synthetic crude oil in the second-stage reactor.

2.1 DOW COAL LIQUEFACTION PROCESS

The Dow Coal Liquefaction process is a catalytic process designed to convert high-sulfur coal into naphtha and low-sulfur fuel oil as liquid products. Ethane/propane mixture (light hydrocarbon gases) and medium-calorific-value fuel gas are produced as salable gaseous products. The byproducts are anhydrous ammonia and elemental sulfur. The process has the following distinguishing features:

- Crushed coal is slurried with recycle oil and liquefied in an upward-flow tubular reactor using patented expendable molybdenum emulsion catalyst in the presence of molecular hydrogen. Liquefaction temperature is 460°C and pressure is kept at about 13.8 MPa.

- The catalyst system is based on emulsion technology and generates a highly active expendable catalyst. The conversion of the catalyst precursor to the active catalyst occurs in the liquefaction reactor. Hydrogen sulfide (H_2S) formed in the conversion of pyrite to pyrrhotite is found to be effective in transformation of molybdenum oxide (MoO_3) catalyst into active molybdenum disulfide (MoS_2). The importance of accessible sulfur for catalyst activation was demonstrated in the recent liquefaction studies carried out with a low sulfur Canadian* coal (0.5 wt% MF coal). Pyrite (FeS_2) added in an amount of 2 kg/100 kg MF coal to ammonium molybdate emulsion [$(NH_4)_6Mo_7O_{24}H_2O$] resulted in sufficient increase of liquid product yields and a decrease in the amount of residual oil (pre-asphaltenes).

---

* These tests with Canadian subbituminous Forestburg coal were carried by Dow Chemical for SNC in 1980-81 as a part of the Technical Assessment of Coal Liquefaction Processes for Canada. The study was done by SNC under a cost-shared contract (No. OSQ79-00090) with CANMET.

- The gases, naphtha and aqueous phase in the slurry leaving the liquefaction reactor are flushed-off and then the remaining slurry is separated by hydroclones and solvent extractor (deasphalter) into fuel oil (deasphalted oil) and a high-solids concentrate suitable as gasifier feedstock to produce hydrogen.

- Hydroclones are used to provide partial solids removal from the product-slurry as well as partial recycle of carried-over catalyst back to the liquefaction reactor. Due to the fine dispersion of the catalyst in the feed, the hydroclone overflow contains about 95% of the catalyst in the original feed.

- A light oil is recovered by liquid-liquid extraction in the deasphalter which is based on counter-current extraction of solids and heavy asphaltenes using a solvent (e.g. hexane).

2.1.1 History and Development

The Dow Coal Liquefaction process has been under development since 1973. The process development studies have been carried out in 'mini-plants' ranging from 1.0 to 110 kilogram-per-day coal feed capacity units. In these studies, catalyst development, solids separation and recycle oil generation were the main objectives. Four patents were awarded on the basis of results obtained. The critical processing steps which were identified include: catalytic hydrogenation of coal in a recycle solvent (deasphalted oil); hydroclones for solid-liquid separation of slurry and, a liquid-liquid extractor so-called deasphalter for producing solids-free coal-derived oil. These processing steps have been demonstrated in a 90-kilogram-per-day, semi-continuous mini-plant.

2.1.1 **History and Development** (cont'd)

In summary, the mini-plant studies established the following typical operating conditions:

Temperature	454°C
Pressure	13.8 MPa
Catalyst	molybdenum trioxide emulsion
Amount of catalyst	250 gram of Mo metal/tonne dry coal
Slurry feed rate	320-640 kg/hr/m^3 reactor volume
Coal in slurry	40 wt%

2.1.2 **Overall Conceptual Plant Configuration**

A simplified flow diagram of the Dow Coal Liquefaction Process is shown in Figure 2-1. Run-of-mine coal is crushed, dried, and pulverized to about 5 mm size and is mixed with process-derived recycle oil to make a 40wt% coal slurry in the slurry preparation block.

The catalyst emulsion is prepared in catalyst preparation block by dissolving molybdenum trioxide in ammonium hydroxide and then emulsifying this solution in the recycle oil. The catalyst emulsion in the amount of 3.3 kg per tonne of slurry is added in the slurry preparation block.

The resultant catalyst-oil-coal slurry is pumped to a short-residence-time preheater, where it is combined with hydrogen and preheated to 420°C. In the preheater, the coal softens and becomes plastic before entering the reactor. The temperature in the reactor is maintained at about 460°C and sufficient residence

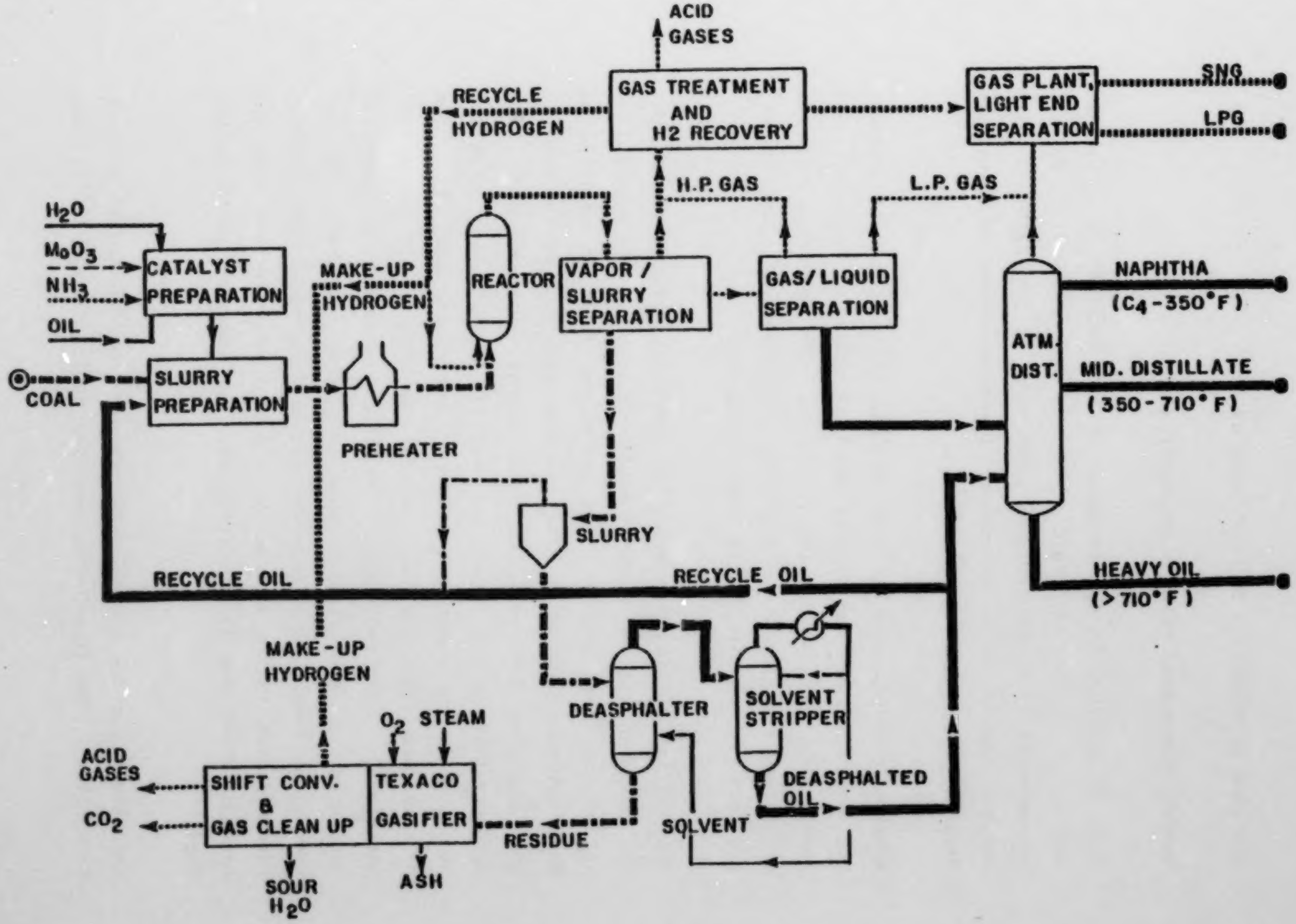

Figure 2-1: Simplified Flow Diagram of Dow Coal Liquefaction Process

2.1.2 **Overall Conceptual Plant Configuration** (cont'd)

time is provided in the presence of the catalyst to allow the more kinetically difficult reactions to proceed. Reactor pressure is 13.8 MPa and coal feed rate is 560 kg/h/m^3. The reactor itself is a back-mixed, pressure vessel of open-tube design.

Reactor products pass through several vapor/liquid separators. The separators are run to take overhead unreacted hydrogen, acid gases, LPG, water vapor, and light oil. The underflow from the separators contains all of the ash, unreacted coal, asphaltenes, and catalyst and most of the distillable oil.

The initial solids separation is performed using hydroclones. The entire hydroclone overhead stream is recycled for preparation of coal/oil feed slurry. The hydroclone split is adjusted so that hydroclone overflow makes up typically 75% of the recycle oil required for coal/oil feed slurry preparation. Ash concentration in the hydroclone overflow is much lower than the ash level in the underflow; however, due to the fine dispersion of the catalyst, the catalyst is not effectively separated. Therefore, the hydroclone overflow contains a catalyst concentration essentially equal to that existing in the hydroclone feed. This catalyst recycle increases the catalyst concentration in the reactor by a factor of 2 or more over the catalyst level due to the fresh catalyst emulsion added to the feed.

Underflow from the hydroclones goes to a countercurrent solvent deasphalter. The deasphalter operates at 2.1 MPa and 180°C. The

2.1.2 **Overall Conceptual Process Configuration** (cont'd)

pressure in the deasphalter is controlled by a back pressure controller and a valve placed on the deasphalter overflow line. In the deasphalter the more polar asphaltenes present in the feed form a second liquid phase which effectively agglomerates the solids in the oil feed. The agglomerated ash, unconverted coal, and asphaltenes are continuously removed from the bottom of the deasphalter column in the form of a very viscous (100 to 200 kg/m s) liquid. The residue from deasphalter is sent to a gasifier for the production of make-up hydrogen.

The overflow from the deasphalter is flashed in the solvent stripper to recover process solvent which is preheated and fed back to the deasphalter. The solvent stripper bottoms stream is an essentially solids-free, low-sulfur product oil. A portion of this oil is used to provide the remaining 25% of the recycle oil required for slurry preparation.

The non-condensed gaseous hydrocarbon streams from vapor/slurry and gas/liquid separation blocks are sent to gas treatment and hydrogen recovery block. The hydrogen is recovered using cryogenic technology. The recovered hydrogen gas (72 wt% H_2, 12 wt% CO and 16 wt% CH_4) is mixed with make-up hydrogen to provide net hydrogen feed for the liquefaction reactor. All non-condensed hydrocarbon gases are treated in the gas plant to recover ethane-propane gas mixture (LPG) and a fuel gas.

The condensed hydrocarbon liquid from gas/liquid separation block along with deasphalted oil from the solvent stripper are fed to an atmospheric distillation column where naphtha, mid-distillate, and heavy distillate oil are produced.

2.1.2 Overall Conceptual Process Configuration (cont'd)

All the acid gas streams are sent to the sulfur plant (not shown in Figure 2-1) to convert H_2S to elemental sulfur.

The plant's sour water streams are collected and treated to produce ammonia and tar acids (water soluble polar aromatics) as byproducts. The ammonia is recovered in a Phosam-W ammonia recovery plant. The effluent water from these units is sent to the water reclamation plant (not shown in Figure 2-1).

The process inputs for a conceptual, commercial plant processing 25000 t/d of as-received Saskatchewan lignites will require about 315 L/s of raw water, and 0.2 kg of molybdenum trioxide (MoO_3) as catalyst per tonne MF lignite. At a cost of $16.80/kg of MoO_3, it amounts to $1.0/barrel of oil equivalent assuming no recovery of spent catalyst.

2.1.3 Technical Review

2.1.3.1 Types of Coal Investigated

Pittsburgh No. 8 (Allison Mine) bituminous coal is the only type of coal extensively tested in the Dow mini-plant units. In 1981, Illinois No. 6 (Burning Star) bituminous coal was processed for three months in a 90-kilogram-per-day computer-controlled mini-plant. Recently, bench-scale investigation of a Canadian Forestburg subbituminous coal from Forestburg, Alberta, has been carried out. Lignites have not been tested in this process configuration.

2.1.3.1 Types of Coal Investigated (cont'd)

In the Dow process, it has been experimentally demonstrated that molybdenum has to be sulfided to become an active catalyst. Testing of Forestburg coal which has a low-sulfur and low-ash content indicated that addition of about 5.6 kg of pyrite per 100 kg of dry coal results in a satisfactory liquefaction response. Since Saskatchewan lignites have higher sulfur content, the addition of pyrite may not be necessary.

2.1.3.2 Product Yields, Distribution, and Efficiencies

Typical product yields and distribution for different coals are presented in the Table 2-1. Also presented in this table is the liquefaction response estimated by SNC for Estevan lignite. The estimates are based on comparative analysis of test results obtained from Forestburg subbituminous coal. The total liquid yields for bituminous coals are about 61 wt% (MAF coal) for Illinois No. 6 coal compared to 52 wt% from Pittsburgh coal. When these values are compared with Forestburg subbituminous coal, the net liquid yield is considerably lower (44 wt%). This is primarily because, Forestburg coal has exceptionally high oxygen content (19 wt% MAF coal vs 10 wt% in Illinois No. 6) and low carbon content. The estimated liquid yields from Estevan lignite is about 32 wt% MAF coal. The low liquid yields from Estevan lignite is because of high oxygen and ash content, and low carbon content.

For Willow Bunch lignite, the decrease in liquid product yield would be even higher than for Estevan lignite due to the higher ash and oxygen content, and low carbon content.

2.1 DOW COAL LIQUEFACTION PROCESS (cont'd)

2.1.3.2 Product Yields, Distribution, and Efficiencies (cont'd)

The estimated overall material balance for a conceptual, commercial, integrated self-sufficient Dow coal liquefaction plant operated with Estevan lignite is presented in Table 2-2 on the basis of 25,000 tonnes of lignite "as-received".

The total liquid yield on moisture and ash free (MAF) basis from Forestburg coal is 7% higher than for Estevan lignite. This is because high ash content in Estevan lignite requires a higher amount of liquid product to be sacrificed for the transport of the residues from the deasphalter.

The overall thermal efficiency for Estevan lignite in the Dow process is estimated as 65.2% based on the net product yields given in Table 2-2. Similarly, the "liquid product" efficiency is 44.5%.

2.1.3.3 Product Quality

The process developer has not made any specific studies pertaining to the analysis of synthetic crude and its upgrading. The analysis of raw naphtha (C_5- 200°C) even for most extensively studied Pittsburgh No. 8 coal has not been reported and therefore, at this stage, no assessment can be made regarding the quality of this product. The analysis of deasphalted oil (200 - 500°C) from this coal however, has been reported. The analysis indicates about 0.5 wt% nitrogen and 2.8 wt% oxygen which is excessively high when compared with similar boiling range material derived from crude petroleum. These impurities will

Table 2-1 Product Yield and Distribution from Different Coals in Dow Coal Liquefaction Process

	Illinois No. 6	Pittsburgh No. 8	Forestburg Alberta	Estevan** Lignite
Process Conditions				
Temperature, °C	455	455	460	-
Pressure, MPa	18.6	13.8	13.8	-
MF Coal Feed Rate, kg/h/m^3	485	480	530	-
Ash in Deasphalter feed, %	40	39	32	-
Product Yields, wt% MAF Coal				
CO_x	1.3	1.4	-	-
H_2S	1.64	2.6	0.93	(1.3)
NH_3 H_2O	11.84	10.7	18.8	(23.4)
C_1- C_3 Gases	13.61	19.6	26.77*	(22.6)*
C_4- 200°C Naphtha	16.53	15.7	13.4	
200-380°C Liquid	22.23	23.5	14.9	(32.0)
>380°C Liquid	21.94	12.5	15.6	
Total Liquid	60.7	51.7	43.9	32.0
Deasphalter Residue	17.64	19.2	15.7	(27.5)
TOTAL	106.73	105.2	106.1	106.8
Hydrogen Consumption	6.73	5.2	6.1	6.8
	100.00	100.0	100.0	100.0
Ash, wt% MF Coal	10.5	8.8	6.1	15.4

* Includes CO_x gases

** Estimated by SNC using Forestburg coal data as the basis

Table 2-2 Estimated Product Yield and Efficiencies in the Dow Conceptual Commercial Plant

[Basis: 25,000 t/d As-Received Coal]

	Forestburg	Estevan*
Coal (As-Received)		
Moisture, wt%	26.1	35.0
Ash, wt% of MF Coal	6.1	15.4
Coal Fed to Reactor		
Moisture, wt%	14.9	14.9
Ash, wt% of MF Coal	6.1	15.4
Input		
MF Coal to Reactor, t/d	18475	16250
MF Coal to Offsites, t/d	-	-
Electricity, MW	96	-
Output, t/d (bbl/d)		
Fuel Gas	1042	870
LPG	952	748
Naphtha (C_5- 200°C)	663 (5880)	569
Light Oil (200 -380°C)	2051 (18192)	2113
Fuel Oil 380°C	4077 (23035)	1690
Sulphur	125	162
Phenol	-	-
Ammonia	45	28
Liquid Product Yield (C_5- 510°C)		
bbl/t As-Received Coal to Reactor	1.9	1.3
bbl/t MF Coal to Reactor	2.5	2.1
bbl/t MAF Coal to Reactor	2.8	2.6
Naphtha in Total Liquid, wt%	51	61
Thermal Efficiency, %		
Overall	68.6	65.2
Liquid Products	50.0	44.5

* Notes For Estevan Lignite

1. Data estimated by SNC.
2. Excess hydrogen (due to excess of oil product in the deasphalter residue to be pumpable) produced by gasification meets entire plant's fuel and power requirements.
3. Product quality and heating value assumed the same as obtained from Forestburg, subbituminous coal.

2.1.3.3 Product Quality (cont'd)

limit the direct utilization of this product. Therefore, product upgrading is required in order to produce marketable end-products.

2.1.3.4 Status of Research and Development

At present, the Dow Chemical Company is operating a 0.1 t/d coal liquefaction 'mini-plant'. Included in this unit are facilities for all the major liquefaction processing steps except coal preparation and residue gasification/hydrogen manufacture. A key feature of this facility is a state-of-the-art computerized process control system. The plant has operated with Illinois No. 6 coal for three months.

The Dow Chemical Company is planning to carry out the following process development activities:

- Scale-up and demonstration of key processing steps: emulsion catalyst system; liquid/solid separation; and deasphalter performance.
- Gasification studies on deasphalter residue followed by molybdenum recovery from the slag.

The work is aimed at defining scale-up parameters and demonstrating commercial viability of the process. The main objective, however, is to develop the process to the extent it becomes ready for licensing.

2.1.3.4 Status of Research and Development (cont'd)

The Dow Chemical Company hopes to utilize existing DOE facilities to scale up its process. Such an effort would likely involve cost-sharing between industry and government. Dow is presently discussing the development of a "Consortium" to utilize the 200-t/d H-Coal facility at Catlettsburg, Kentucky, to scale up its catalyst and deashing systems. Dow is also negotiating with other companies and U.S. Department of Energy as an alternative to scale-up some of the processing steps using the existing facilities.

In order to reach the readiness for commercialization, the following R & D work will be required:

- Demonstration of process suitability with different rank coals particularly lignitic coals which have high ash and moisture contents.
- Implementation of a sound experimental program on a pilot-plant scale to provide scale-up data for engineering design of a conceptual commercial integrated plant.
- Examine potential problem areas, process reliability and selection of proper materials of construction.
- Investigation of product properties and upgrading routes for the disposition of synthetic crude as marketable synthetic fuels or petroleum feedstocks.

Recognizing the magnitude of efforts required for the development of the Dow coal liquefaction process, it appears that this process will not reach commercialization readiness before 1995.

2.1.3.4 Status of Research and Development (cont'd)

At present, there are no long-range commitments either from process developer or from the U.S. Government. This may further delay the commercialization prospects of the process.

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2.2 CO-STEAM PROCESS

The CO-Steam process is a second generation coal liquefaction technology being developed for converting low-rank coals to environmentally acceptable syncrude. Further fractionation or hydrotreating of the syncrude product may result in production of distillate products such as heating oil, gasoline,fuel oil and/or chemical feedstocks. The process concept utilizes the reaction of carbon monoxide (or synthesis gas) and steam with or without addition of catalyst. The process is specifically designed for lignites with high reactivity and high moisture content.

The main feature of the process concept is the reaction of carbon monoxide (CO) and steam (resulting from existing moisture in coal) with coal which provides in-situ hydrogen for liquefaction. The CO also removes oxygen from coal as CO_2 thus reducing hydrogen consumption.

2.2.1 History and Development

The conversion of coal to oil by reaction with carbon monoxide (CO) and steam was first reported in 1921 by Fischer and Schrader. This work was extended in batch autoclave studies at Pittsburgh Energy Research Center (PERC) beginning 1966. In 1975, the U.S. Department of Energy (DOE) assigned its Grand Forks Energy Technology Center (GFETC) in North Dakota to continue further development work on the concept of the CO-Steam process. At this stage, the main concern was related to high product viscosity, high pressure (27.6 MPa), and an excessively long reaction time of 1 to 2 hours as observed in a series of experiments carried out at PERC. Thus, the main objectives at GFETC were to develop the chemical and engineering data base required to liquefy lignites and subbituminous coals under moderate conditions.

The liquefaction research facilities at GFETC include a hot-charge and time-sampled batch autoclave system and a 2.3 kg/h continuous process unit (CPU). The CPU has two alternate reactor arrangements, continuous and stirred. Design and construction of CPU started in 1975. This facility has been designed to provide flexibility for changing backflow (residence time distribution) and level of turbulence to permit the systematic study of improvements in reactor design. Supporting studies in the batch system involved determination of reaction kinetics and extensive work on product characterization using an array of analytical instrumentation.

2.2.1.1 Batch Autoclave Studies

Batch autoclave studies at GFETC were carried out in a 1-litre magnetically stirred autoclave. A schematic of time-sampled batch autoclave and an on-line gas chromatograph system employed in the studies are shown in Figure 2-2. The unit is outfitted for rapid charging of slurry to the preheated reactor and for time sampling of gas and slurry during experiments. The other important feature of this unit was an on-line gas chromatograph system.

The main process variables studied are the effects of reactant gas composition, the solvent (tetralin-anthracene oil mixture), and the temperature on liquefaction yields. The ranges of pressure and temperature investigated are from 20.7 to 33.1 MPa and 350 to 480°C respectively.

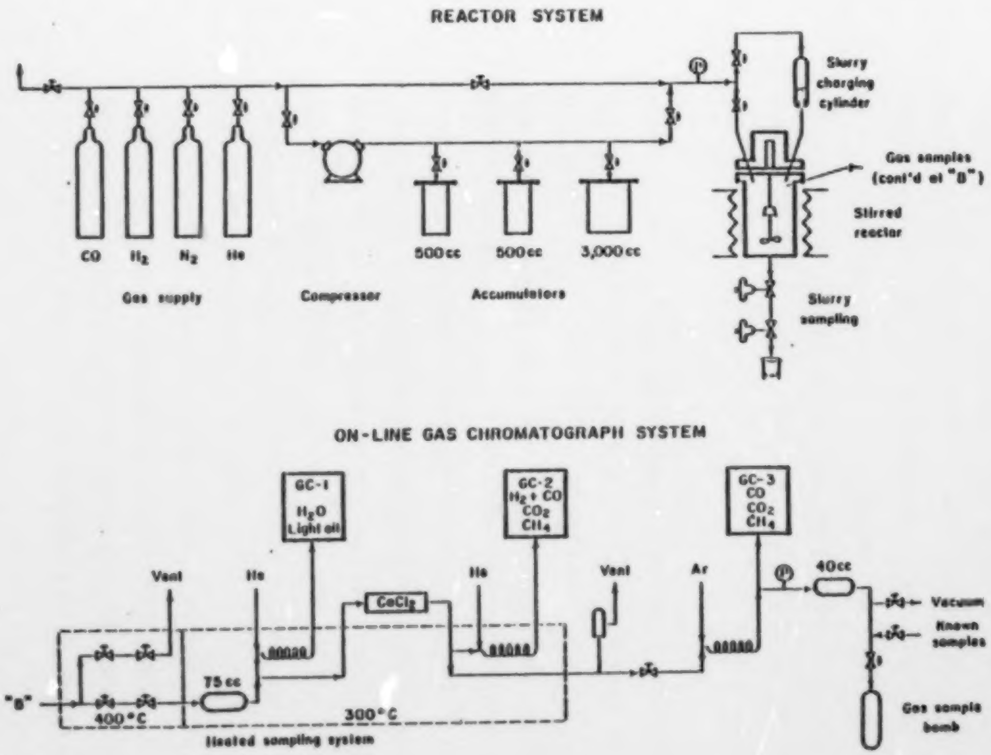

Figure 2-2: Schematic Diagram of the time-sampled batch autoclave and on-line gas chromatograph system

Source: Sondreal, E.A., C.L. Knudson, J.E. Schiller, and T.H. May. Development of the CO-Steam Process for Liquefaction of Lignite and Western Subbituminous Coals, Proceedings, 1977 Symposium on Technology and Use of Lignite, Grand Forks, ND, May 1977; DOE GFERC/IC-77/1, pp. 129-158.

2.2 CO-STEAM PROCESS (cont'd)

2.2.1.1 Batch Autoclave Studies (cont'd)

The results of process variable studies indicated that CO-steam liquefaction of lignites involves two categories of reactions: (1) reactions occuring over the entire temperature range (350-480°C) which through rapid consumption of CO produce major portion of the total yield of product oil (either soluble oil or distillate oil), and (2) reactions occuring at or beyond 460°C and involves depolymerization of the heavy ends (high molecular weight fractions).

The effect of residence time on liquefaction yields using H_2/CO mole ratio of 1 in syngas at 437°C and 27.6 MPa (4000 psig) is shown in Figure 2-3. The conversion (wt% MAF coal basis) of lignite to gas and soluble product (A + B + C) is very rapid and reaches 70% in 10 minutes, followed by a slow increase to 90% after about 2 hours. The gas yield comprised oxygen removed from coal and appearing as CO_2, along with smaller amounts of methane, ethane, and propane. This gas yield approaches 20% in 30 minutes and increases thereafter to about 30%. The yield of soluble oil (B + C) reaches 49% in 10 minutes and thereafter increases slowly to about 60%; the non-distillable portion (C) determines by its quantity and quality (whether the total soluble product is a liquid or a solid). Higher reaction temperatures tend to reduce both the quantity of the "C" fraction and its average molecular weight, with an increase in distillable oil (B) and gas (A).

2.2.1.2 Bench-Scale Program

During the development of the CO-Steam process, extensive use of 2.3 kg/h bench-scale unit called 'Continuous Process Unit' (CPU) has been made. This facility is located at GFETC and a schematic flow diagram of this unit is shown in Figure 2-4.

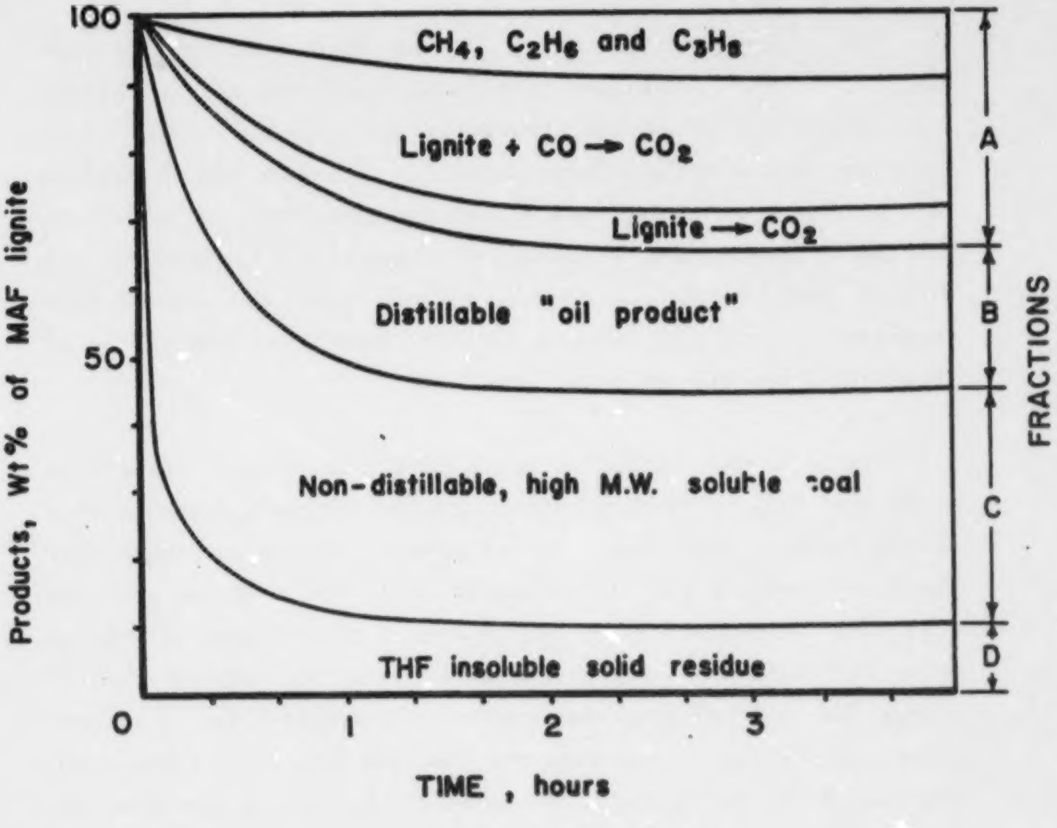

Figure 2-3 Effect of Residence Time on Liquefaction Yields

Source: E.A. Sondreal, C.L. Knudson, J.E. Schiller, and T.H. May. Development of the CO-Steam Process for Liquefaction of Lignite and Western Subbituminous Coals, Proceedings, 1977 Symposium on Technology and Use of Lignite, Grand Forks, ND, May 1977; DOE GFERC/IC-77/1, pp. 129-158.

The CPU system consists of : (1) an input section including coal/oil slurry preparation tank (not shown) and slurry holding tank along with slurry recirculation and injection pumps, feed gas compressor and a gas surge tank; (2) alternate slurry preheaters (stirred or tubular) and a feed gas preheater; (3) alternate reactors (tubular and stirred); (4) gas/liquid separators and recycle loop (including slurry recycle pump and gas recycle compressor) and; (5) output section including heavy product receiver, light oil and water receiver.

Pulverized lignite (minus 60-mesh) having about 30% moisture is preblended with solvent or recycle product slurry and pumped to a slurry holding tank, which is stirred by slurry recirculation. The slurry holding tank is suspended on a load cell for accurate feed rate determination. A high-pressure piston pump delivering up to 11.3 kg/h of slurry at pressures up to 41.4 MPa is used to charge the slurry into the system. A single-stage diaphragm compressor is used to compress the reaction gas to 43.4 MPa. The compressed gas is regulated and metered into the system from the surge tank. Slurry and gas are heated independently and mixed at the entrance of the reactor.

The stirred reactor is a 3.8 litre magnetically-stirred autoclave with two marine-type propeller stirrers circulating the slurry downward. The autoclave has a total volume of 3785 cm^3, but a dip tube limits slurry volume to about 2458 cm^3. At a normal slurry flow rate of 4.5 kg/h, the average residence time of the liquid is approximately 30 minutes. The reactor, constructed of type 316 stainless steel, rated at 34.5 MPa at 500°C. Normal operating pressure is 27.6 MPa. Slurry can be sampled directly from the reactor through a two-valve system which removes 3 ml per cycle.

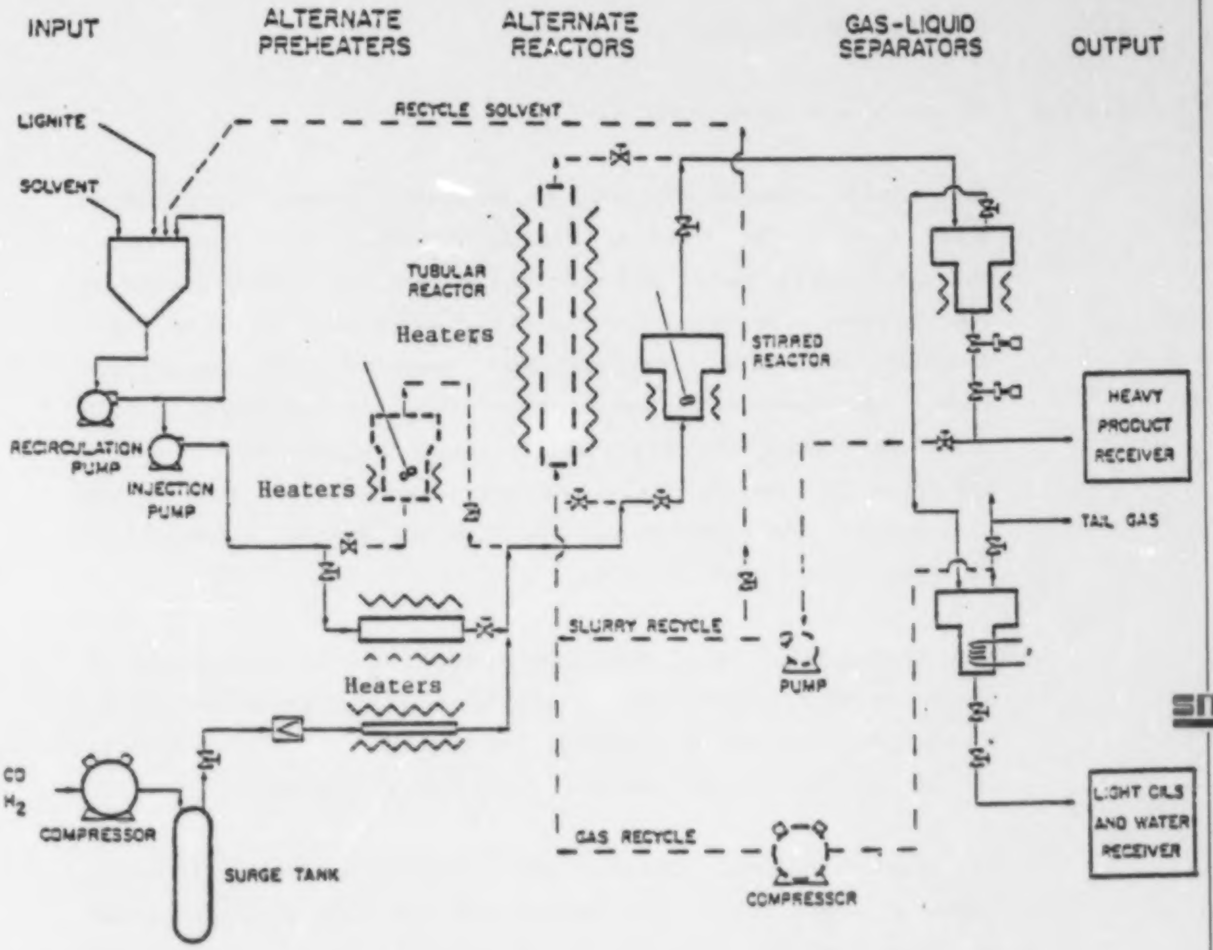

Figure 2-4: Schematic diagram of the GFETC 2.3 kg/h continuous process unit

Source: W.G. Wilson, C.L. Knudson, G.G. Baker, T.C. Owens, D.E. Severson "Application of Liquefaction Processes to Low Rank Coals". Proceedings of the Tenth Biennial Lignite Symposium Sponsored by U.S. Department of Energy. May 30-31, 1979 at Grand Forks, N.D. GFETC/1C-79/1 (Conf-790579)

2.2.1.2 Bench-Scale Program (cont'd)

The tubular reactor employed is made of 1.5-meter long, 38-mm steel tube. The reducing gas is injected into the reactor through a porous metal diffuser built into the bottom flange of the reactor. Reactor temperature is maintained by five individually controlled external-jacket heaters. The temperature inside the reactor is measured by five thermocouples - four introduced through the side wall, starting 30.5 cm off the bottom and up at 30.5 cm intervals, and one extending just through the top flange. The temperature rise from the bottom to the top of the reactor is typically 40°C.

The objectives of using alternate reactors are to investigate if the rates of reaction could be enhanced by increasing turbulance through stirring and by changing the residence time distribution (in the tubular reactor compared to a stirred reactor).

The reactor effluent enters a high-pressure hot separator maintained at about 300°C. The liquid level in this pressure vessel is controlled by a nuclear level detector which actuates the sequential operation of two valves. These pressure let-down valves release 25 ml of slurry every 25 seconds into a heavy product receiver.

Gases and vapors from the hot separator pass to a condensate accumulator where the light oils and water are condensed by water-cooled coil. The liquid level in this vessel is monitored by a nuclear level detector. Light oils are removed from the system through a series of two valves which are operated manually. The tail gas is let down through a backpressure regulator and metered and analyzed using an on-line gas chromatograph.

2.2 CO-STEAM PROCESS (cont'd)

2.2.1.2 Bench-Scale Program (cont'd)

The CPU was operated under a "once-through" mode using a 40% to 60% blend solvent from two stock materials: the first designated as SRCMD was a mid-distillate fraction produced in the pilot plant operating in SRC-II mode at Fort Lewis, Washington; the second solvent was anthracene oil (AO4) obtained from the Crowly Tar and Chemical company. Both of these solvents had low coking properties and low hydrocarbon gas production.

A series of CPU runs were also made under "bottoms recycle operation mode". Recycle experiments were carried out by pumping feed slurry through the system and collecting the product slurry over 3- or 4-hour intervals for subsequent recycle. The collected product slurry was sampled, weighed, and a portion blended as the pasting solvent with "as-received" pulverized lignite. The resulting mixture was either recycled back to the reactor as prepared or first partially dried in an electrically-heated drying tank to arrive at the desired feed moisture content. Since the slurry blend tank and feed tank are maintained at less than 100°C, it is usually necessary to add back some of the light distillates (representing up to 5 wt% of the slurry fed to the reactor) for feed slurry dilution and viscosity control.

The CPU was mainly operated to carry out process variables studies. The results of process variables study are as follows:

- Increasing coal concentration in the feed slurry from 30 wt% to 48 wt% does not result in significant change in liquefaction yields.

2.2.1.2 Bench-Scale Program (cont'd)

- When only anthracene oil is used to prepare coal slurry, the net liquid yields (expressed as percentage solubility in tetrahydrofuran, THF) decrease with increasing temperature from 440 to 480°C. However, the yields increase with temperature when solvent consists of 7 wt% tetralin in the anthracene oil. (These experiments were performed with 40% coal slurries in a continuous flow stirred reactor, at a nominal pressure of 27.6 MPa and using syngas containing 1:1 mole ratio of H_2 to CO).

- Increasing the temperature from 440 to 480°C, the distillate product yields increase from 30 wt% to about 62 wt% MAF coal when the solvent contained 7 wt% tetralin in anthracene oil. Similarly the C_1-C_4 hydrocarbon gases yield also increases from 6 wt% to 22 wt%. However, the gas yields were higher by about 10 wt% when the solvent did not contain tetralin. Thus donor solvent was found to enhance liquid yields while reducing gas yields.

- The distillate yields increase from about 31 wt% (once-through operation) to 46 wt% MAF coal under "bottoms recycle" operation.

- Based on the differences in molecular weight distribution in the products as a function of temperature, it was found that the stirred reactor would be less effective than a tubular reactor with adequate radial gas-slurry contact.

2.2.2 Technical Review

2.2.2.1 Types of Coal Investigated

North Dakota lignites from 'South Beulah' and 'Indian Head' mines are the only coals extensively tested in both batch autoclaves and bench-scale (CPU) unit. Some testing has also been done with Texas 'Big Brown' lignite. The ash and moisture in North Dakota lignites tested were in the range of 7 wt% to 12 wt% ash and 28 wt% to 37 wt% moisture. Proximate and ultimate analysis of these lignites are presented in Appendix A.

The Willow Bunch lignite from Saskatchewan has similar analyses as the South Beulah and North Dakota lignites. On MF basis, the volatile matter is practically the same (40.6 wt% vs 39.9 wt%). The fixed carbon is about 4 wt% less and the ash content is 4 wt% more in Willow Bunch when compared with South Beulah lignite.

2.2.2.2 Product Yields and Distribution

Typical net product yields (expressed as wt% MAF coal) from North Dakota lignite under "once-through" and "slurry-recycle" modes of operation are presented in Table 2-3. It can be seen that the overall conversion on MAF basis in an tubular reactor is higher than in a continuous stirred reactor under both, once-through and slurry-recycle modes of operation.

2.2 CO-STEAM PROCESS (cont'd)

2.2.2.2 Product Yields and Distribution (cont'd)

A comparison of liquid yields (wt% MAF coal) indicating the effect of moisture, pressure and use of syngas over hydrogen is presented in Table 2-4. These results were obtained under "bottoms-recycle" operation in the bench-scale unit (CPU).

When lignite is fed 'as-received' containing 30 wt% moisture, the total liquid yield is 36 wt% with syngas compared to 61 wt% with hydrogen under same liquefaction conditions (13.8 MPa and 460°C). The yields increase to 71 wt% with syngas if the pressure is increased to 27.6 MPa. However, lowering the pressure to 13.8 MPa and also reducing the moisture content of feed coal to 17 wt%, the total liquid yields of 67 wt% is achieved with syngas (coal with moisture content below 17 wt% could not be tested with the existing slurry drying equipment).

The estimated reactor yields from Willow Bunch lignite would be similar to those obtained from Beulah lignite.

2.2.2.3 Product Quality

The liquid produced from the CO-Steam liquefaction of lignite consists mainly of distillable oil and about 15 wt% to 20 wt% (MAF coal) of non-distillable liquid residue which is tetrahydrofuran (THF) soluble material. The non-distillable portion has been found to contain asphaltenes and preasphaltenes. The chemical analysis and characterization of these liquids are not available.

Table 2-3 Typical Bench-Scale Yields in CO-Steam Process

	Once-Through		Bottom-Recycle	
Reactor Type	OTR[a]	CSTR[b]	OTR[a]	CSTR[b]
Experimental Conditions				
Temperature, °C	465	461	460	460
Pressure, MPa	14.1	27.2	27.6	27.6
Reducing Gas	H_2	H_2+CO	H_2+CO	H_2+CO
Solvent	SCRMD[c]	AOD[d]		
Net Yields, wt% MAF Coal				
CO_2	7.4	22.6[e]	11.0[f]	26.1[f]
H_2S + NH_3	2.7	-	-	0.5
H_2O	11.5	-22.4	-	-
CH_4	2.0	2.8	5.0	12.9
C_2 - C_4	3.3	3.4	10.6	12.1
C_4 + Liquid	38.5	41.2	51.1	21.8
Residue	34.1	31.7	20.4	18.3
Insoluble Organic Matter (IOM)	9.9	22.8	7.9	15.1
	104.9	102.2	106.0	106.8
Overall Conversion, % MAF Coal	90.1	77.2	92.1	84.9

a: Open tube reactor

b: Continuous stirred tank reactor

c: Blend of 40% hydrogenated anthracene oil and 60% middle distillate fraction from SRC-II pilot plant at Ft. Lewis, Washington

d: Anthracene oil distillate

e: Includes H_2S and NH_3

f: Includes H_2O

Source: W.G. Wilson et al., "Comparison of Liquefaction Yields", 1981 Lignite Symposium, San Antonio, Texas, June 15-17, 1981.

Table 2-4 Comparison of Liquid Product Yields under Different Conditions in Co-Steam Process

	North Dakota 'Beulah-3'				Texas 'Big Brown'
Liquefaction Conditions					
Coal Moisture, wt%	30	30	30	17	26
Temperature, °C	460	460	460	460	460
Pressure, MPa	13.8	13.8	27.6	13.8	17.9
Reducing Gas	$CO+H_2$	H_2	$CO+H_2$	$CO+H_2$	H_2
Liquid Yields, wt% MAF Coal					
Distillate Oil	22	45	51	46	46
Non-Distillate Liquid	14	16	20	21	17
Total Liquid	36*	61	71	67	63
Distillate - to - Non Distillate Ratio	1.6:1	2.8:1	2.6:1	2.2:1	2.7:1

Notes: (i) "bottoms-recycle " operation

(ii) "bench-scale results

* Lower total liquid yields resulted when catalytically hydrogenated anthracene-oil bottoms were used as solvent to prepare coal slurry. In all other cases, coal-derived light oil was used.

Source: W.G. Wilson et al., "Comparison of Liquefaction Yields", 1981 Lignite Symposium, San Antonio, Texas, June 15-17, 1981.

2.2.2.4 Status of Research and Development

To date, the process concept has not been optimized for pertinent operating parameters, nor is the mechanism known for CO reactions with water for supression of carbonate or coke formation in the reactor. A great deal of R & D work is still required before the process concept could be judged towards commercialization.

However, at present no further R & D work on CO-Steam concept is planned at Grand Forks Energy Research Center because of cuts in funding.

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2.3 CONSOL SYNTHETIC FUEL PROCESS

The Consol Synthetic Fuel (CSF) process is one of the first two-stage coal liquefaction processes being developed in the U.S.A. The Office of Coal Research (predecessor of DOE) sponsored the construction of a multipurpose liquefaction test center at Cresap, West Virginia for research and development activities utilizing the CSF process. In the first stage of this process, coal is extracted in a coal-derived solvent. After separation of mineral matter, hydrogenation of extract is carried out in the second stage to yield synthetic crude.

The features and key processing areas of the CSF process include:

- Coal extraction at low pressure between 1.0 to 2.7 MPa (150 to 400 psig) in the presence of a hydrogen donor solvent.
- Liquid/solids separation by solvent deashing.
- Low temperature (482°C, 0.06 MPa) carbonization of solid-rich slurry to produce additional liquid, char and to recover deashing solvent.
- Catalytic hydrogenation of coal extract and carbonization liquid at 20.7 MPa and 427°C to yield donor solvent, raw naphtha and fuel oil.
- Gasification of char to produce hydrogen.

2.3 CONSOL SYNTHETIC FUEL PROCESS (cont'd)

2.3.1 History and Development

The Consolidation Coal Company* developed the CSF process using a 4.5 kg/h bench-scale unit at Library, Pennsylvania in the early 1960's. On the basis of this research the Office of Coal Research sponsored the construction and operation of a 18-t/d plant at Cresap, West Virginia. The pilot plant began operation in November 1966, and was terminated in February 1970. The plant operation had several mechanical problems. However, the nature of these problems has been withheld by the process developer.

In 1974, Fluor was contracted by the Office of Coal Research (later to become ERDA and currently DOE) to refurnish the Cresap facility into a multi-purpose coal liquefaction test center and conduct development work, utilizing the CSF process concept. During 1974-1977, the plant was successfully revamped. In 1977 the plant operations started. Experimental program including numerous tests were carried out from 1977 till September 1979, when the project was terminated.

A simplified flow diagram of the CSF 4.5-kg/h coal feed capacity bench-scale unit at Library, Pa., is shown in Figure 2-5. The unit consists of slurry feeder, extractor, solids separation system, and a fluidized-bed hydrocracker reactor (not shown in the figure).

Dried and pulverized coal (208 μm) is premixed in 15 kg batches using extraction solvent in the ratio of 2 to 1 and then is pumped into the preheater-extractor.

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* Now the CONOCO Development Company, a subsidiary of the Continental Oil Company.

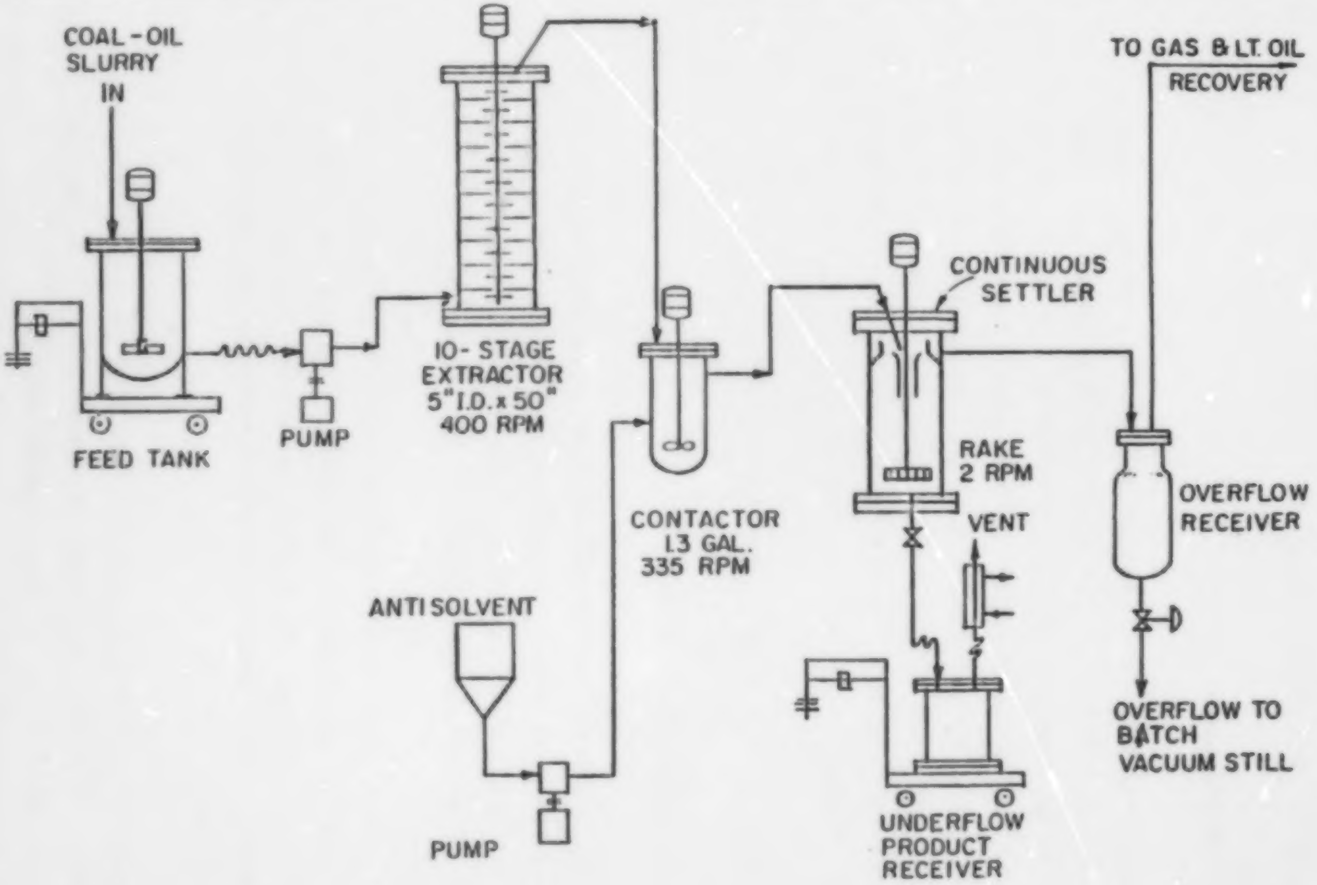

Figure 2-5: Coal Extraction and Liquid-Solid Separation Systems of 4.5 kg/h CSF Bench-Scale Unit.

Source: DOE Report FE-1517-T68, November 1978.

2.3.1 **History and Development** (cont'd)

The preheater-extractor is a 10-stage continuously stirred vessel with a volume of 0.016 m^3. The slurry in each "stage" is mixed by turbine blade impeller baffles rotating at 375 rpm. In the five "bottom stages", the slurry is preheated to the extraction temperature, while in the remaining five "top stages" serve as the extractor where the slurry is kept under turbulent regime and temperature and holding time corresponding to designed conditions. Typical extraction conditions are:

- Temperature : 326 to 400°C
- Pressure : 28 to 4 MPa
- Solvent-to-Coal Ratio: 1-5 to 2-0

The extractor effluent is transferred to a 5-litre contactor where anti-solvent is added to promote separation of the ash, heavy extract and unreacted solids. The contactor overflow is transferred to the continuous settler.

The heavy material and solid residues (settler underflow) are withdrawn manually at 15-minute interval. The gas and liquid stream (settler overflow) is further separated into gases and extract streams. The extract produced is sent to the catalytic hydrogenation unit.

Typical yields from extraction and hydrocracking stages are shown in Table 2-5. The coal feedstock is Pittsburgh No. 8 and the extraction is done with a commercially available solvent - "Sure-Sol" blended with 12 wt% tetralin. The boiling range of Sure-Sol is 207 to 366°C.

Table 2-5 Typical Reactor Yields in the CSF Process

	Extraction Stage	Hydrocracking Stage	Combined
Operating Conditions			
Temperature, °C	400	454	
Pressure, MPa	3.8	24.1	
Solvent-to-Coal ratio	1.78	-	
Yields, wt% MAF Coal			
CO_x	0.4	0.4	0.8
H_2S	1.7	1.6	3.3
$NH_3 + H_2O$	0.3	5.0	5.3
$C_1 - C_4$	0.8	12.1	12.9
C_5 - 300°C	9.9	12.7	22.6
300° - 470°C	70.7	28.8	28.8
470°C		15.2	15.2
MAF Residue	16.1	-	16.1
	100.0	75.8	105.0
H_2 Consumption	0	5.0	5.0
	100.0	70.8	100.0

Notes: (i) Extraction stage conversion 83.9 wt%

(ii) Hydrocracking stage conversion 78.8 wt%

(iii) Hydrocracking yields expressed as wt%

Sources: DOE Report FE-1517-T68, Nov. 1978.

2.3.2 Overall Conceptual Plant Configuration

In 1976, the Office of Coal Research awarded a contract to Fluor Engineers and Constructors to carry out a conceptual design of a commercial plant (50,000 barrels per day of liquid products) based on the Consol Synthetic Fuel (CSF) process. A simplified flowchart of main liquid and residue (char) streams based on this conceptual design is shown in Figure 2-6.

A mixture of recycle donor solvent and non-hydrogenated solvent is used for preparation of coal slurry. Atmospheric extraction at rather high temperature (404°C) is applied for extraction of coal into liquid followed by antisolvent deashing. Settler's overflow, i.e., extract after solvent recovery, is sent to hydrogenation in an ebullated-bed catalytic reactor operating at severe conditions (20.7 MPa pressure and 427°C temperature).

After gas/vapor - liquid separation of reactor effluent, the raw liquid product (syncrude) is fractionated into raw naphtha, donor solvent, fuel oil and vacuum residue streams. Raw naphtha and fuel oil are sent to product refining (upgrading) to gasoline. Donor solvent is recycled back for coal slurry preparation in a mixture with non-hydrogenated solvent. Vacuum residue is used in an oxygen blown gasifier for production of make-up hydrogen.

Settler underflow, containing mineral matter (ash), unreacted coal and solvent undergoes low temperature carbonization in the fluidized-bed carbonizer in order to produce char, additional liquid hydrocarbons (tar) and to recover solvent.

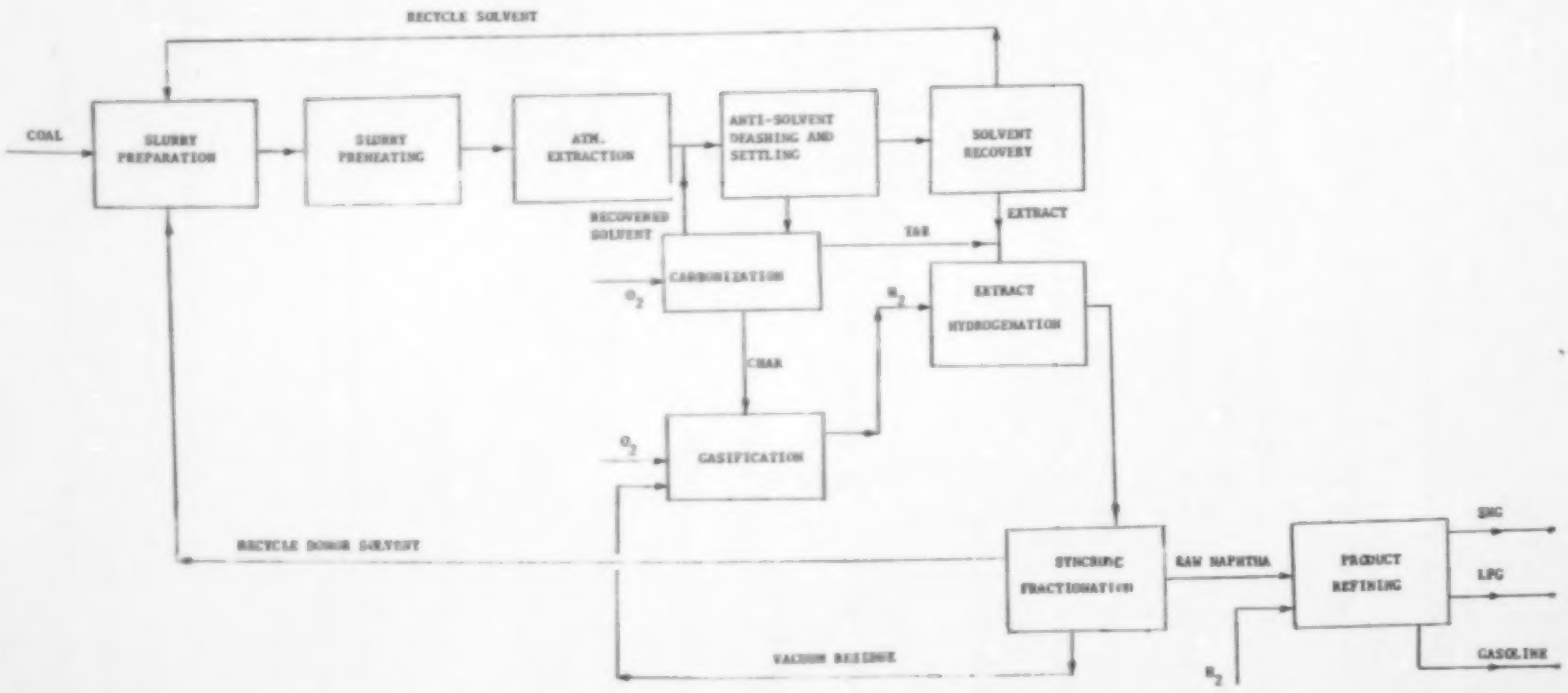

Figure 2-6: Simplified Flow Diagram of Major Liquid and Residue Sreams in the CSF Process

2.3 CONSOL SYNTHETIC FUEL PROCESS (cont'd)

2.3.2 **Overall Conceptual Plant Configuration** (cont'd)

Recovered solvent is pumped back to slurry deashing while tar is sent for hydrogenation. Char is gasified together with vacuum residue to produce make-up hydrogen.

The ancillary facilities to provide an environmentally acceptable plant operation and recovery of corresponding byproducts (sulfur, ammonia, phenols, etc.,) were included in this conceptual design (not shown in Figure 2-6).

Product yield and efficiencies of the plant estimated in this conceptual design are presented in the following section.

2.3.3 **Technical Review**

2.3.3.1 Types of Coal Investigated

In the bench scale unit and pilot plant, Pittsburgh No. 8 bituminous coal from the Ireland mine has been tested only. This coal contained about 9 wt% ash and only 1.2 wt% moisture.

2.3.3.2 Product Yields and Efficiencies

Typical extraction reactor yields, expressed as wt% MAF coal, was presented earlier in Table 2-5. The subsequent hydrocraking reactor yields were also included.

Product yields and distribution of a conceptual, commercial CSF plant operating with Pittsburgh No. 8 bituminous coal are shown in Table 2-6. The data are based on the conceptual plant data of CSF process prepared by Fluor Engineers and Constructors.

Table 2-6 Estimated Product Yield and Efficiencies in the CSF Conceptual Plant

[Basis: 25,000 t/d As-Received Coal]

	Pittsburgh No. 8
Coal (As-Received)	
Moisture, wt%	1.01
Ash, wt% of MF Coal	10.0
Coal Fed to Reactor	
Moisture, wt%	1.01
Ash, wt% of MF Coal	10.0
Input	
MF Coal to Reactor, t/d	24747
MF Coal to Offsites, t/d	-
Electricity, MW	-
Output, t/d (bbl/d)	
SNG	1872
Propane	527
Butane	656
Gasoline (53.6° API, 89 RON)	7297 (54508)
Sulfur	1109
Phenol	29
Ammonia	210
Liquid Product Yield (gasoline)	
bbl/t As-Received Coal to Reactor	2.18
bbl/t MF Coal to Reactor	2.20
bbl/t MAF Coal to Reactor	2.44
Thermal Efficiency, %	
Overall	62.6
Liquid Product* (gasoline)	42.6

Source: DOE Report FE-2251-40, Vol. 1, 2, 3, October 1977.

2.3.3.2 Product Yields, Distribution and Efficiencies (cont'd)

The liquid product is 89 RON clear gasoline and the yield is 32.95 wt% MF coal or 32.8 wt% MAF coal.

The estimated overall and liquid product efficiencies for a commercial CSF plant operating with Pittsburgh No. 8 coal are 62.6% and 42.4%, respectively.

In terms of availability of process inputs, CSF conceptual commercial plant will require about 100 t/d of make-up antisolvent (e.g., Soltrol 130 - Phillips Petroleum Company) and between 606 to 883 L/s of raw water from an outside source.

2.3.3.3 Product Quality

The process developer has not caried out any specific studies pertaining to liquid product properties and their upgrading with respect to end-use compatibility. However, elemental analysis of primary distillate products are presented in Table 2-7. The elemental analysis reveals that the raw liquid products contain higher concentration of nitrogen and oxygen heteroatoms, as compared to conventional fuel oil. Therefore these distillates or synthetic crude produced will require further hydrotreating.

2.3.3.4 Status of Research and Development

The multi-purpose coal liquefaction test unit (18 t/d designed throughput) based on a two-stage CSF process concept was constructed in 1974-1977 at Cresap, W. Va. under the sponsorship

Table 2-7 Elemental Analysis of CSF Process Distillate Streams

	Elemental Analysis, %				
	C	H	O	N	S
Light-Distillate (C_5 - 180°C)	85.83	12.4	1.3	0.4	0.07
Mid-Distillate (180 - 230°C)	87.7	10.7	1.1	0.4	0.1
Heavy-Distillate (> 230°C)	87.86	9.2	2.5	0.4	0.04

Source: DOE Report, FE-1517-T68, November 1978.

2.3.3.4 Status of Research and Development (cont'd)

of the U.S. DOE. Renovation of this center involved modification of existing facilities to meet environmental requirements and current design standards of piping, electrical, and mechanical equipment codes. A task force was established to organize, coordinate, and direct the testing program.

From late 1977 to September 1, 1979, the coal liquefaction test center at Cresap was processing low-moisture and low-ash Pittsburgh No. 8 bituminous coal. All the major processing steps (coal extraction, deashing and hydrogenation of ash-free extract) were tested. More than 350 hours of operation (including 40 hours fully integrated operation) were carried out. Many problems were encountered during these tests such as plugging of slurry preheater, failure of shaft seal in extractor, blockage of deashing system and others.

On September 1, 1979, the CSF facilities at Cresap and the CSF project as a whole was terminated.

At present, no further R&D work on CSF process or use of Cresap test facilities is considered, either by U.S. DOE or private industry.

Nevertheless, the experience gained in the course of CSF pilot plant operation has contributed to the advances in coal liquefaction technology. The TSL process being developed by C-E Lummus and Cities Service may be seen as an example of further development in two-stage liquefaction concept utilizing the CSF pilot plant experience at Cresap.

2.4 TWO-STAGE LIQUEFACTION PROCESS

The integrated Two-Stage Liquefaction (TSL) process is a second generation coal liquefaction process being jointly developed by the Lummus Company and Cities Service Research and Development Company. TSL is a combination of Short Contact Time (SCT) coal hydroliquefaction as the first-stage, and hydrotreatment of ash-free coal extract in ebullated-bed reactor called LC-Fining as the second stage. The mineral matter (ash) is removed from the extract before it enters the LC-Fining using proprietary Lummus Antisolvent Deashing Technology (ASDA). The features of the TSL process are:

- The SCT hydroliquefaction step (first stage) is operated at conditions to produce high conversion of coal to liquid, but the reaction need not be in solvent balance nor make a marketable product. The reaction conditions are thereby moderated to consume less hydrogen and make fewer hydrocarbon gases (C_1-C_4).

- The LC-Fining (second stage) is run at less severe hydrocracking conditions than would be required for a one-stage coal hydroliquefaction process.

- The feed to the second stage is a deashed extract so that the catalyst undergoes less severe service than in a reactor containing unreacted coal and ash.

- The upgrading reaction (second-stage) can be operated over a range of conditions to provide different product distributions.

2.4 TWO-STAGE LIQUEFACTION PROCESS (cont'd)

- The upgrading stage provides solvent for the liquefaction stage. The recycle solvent may be selected from a number of flash or distillation operations so that the boiling range of the recycle solvent may be varied to optimize the liquefaction operation.

2.4.1 History and Development

The development and demonstration of the TSL process was carried out after the anti-solvent deashing (ASDA) and L-C Fining technologies had been developed independently. Therefore, in the review of TSL process, the development history of ASDA and L-C Fining processes are presented.

2.4.1.1 Antisolvent Deashing (ASDA)

The development of the Antisolvent Deashing process began in 1971 at the C-E Lummus Engineering Development Center in Bloomfield, N.J. The process employs a promoter liquid or antisolvent whose chemical nature results in agglomeration of the micron-sized particles present in the liquefied coal solution. Because of agglomeration, separation of a clean product containing less than 0.2 wt% ash can be achieved by gravity settling of particles. A key variable in the deashing process is the paraffinicity of the antisolvent as measured by its Watson characterization factor (for details see U.S. Patents 3,852,182 and 3,856,675). Initial demonstration of the ASDA unit was carried out at the Lummus ASDA pilot plant. The material processed was a topped filter feedstock from an SRC pilot plant operated by the Pittsburg and Midway Coal Mining Co. at Fort Lewis, Washington. This feedstock, which contained 61 wt% solvent, 31 wt% SRC, and 8 wt% ash and unconverted coal, was deashed during 121-hour pilot plant run.

2.4.1.1 Antisolvent Deashing (ASDA) (cont'd)

The results indicated that:

- A recovery between 75% to 80% of the SRC product present in the feedstock is achieved.
- The ash-rich underflow generated during the test could be vacuum stripped to a 370°C cut point without loss of fluidity.
- No buildup of carbonaceous deposits had occurred on the internal surface of the gravity settler.

This work led to design and construction of ASDA at the 45-t/d SRC pilot plant at Fort Lewis under contract from the U.S. Department of Energy (DOE). The C-E Lummus Company was responsible for basic and detailed engineering as well as procurement and construction supervision work which began in March 1977. The unit was put into operation during the early part of 1979. The design basis for this ASDA unit was developed through onsite batch testing at Fort Lewis using the process information obtained at the bench-scale deashing unit in Bloomfield (antisolvent-to-feed ratio, antisolvent characterization factor, settler residence times and velocities, etc.). Details of this work are proprietary. The design of the Fort Lewis ASDA unit included a number of experimental features for the purpose of optimization studies such as: (1) alternate settler feed locations to demonstrate the effect of decreased underflow residence time; (2) a method to determine the maximum process capacity of a single settling compartment, etc.

2.4.1.1 Antisolvent Deashing (ASDA) (cont'd)

The unit was operated for 4 months and SRC recovery of up to 75% was achieved. In the ash residue, the pyridine solubles level of about 50% was found necessary to keep the residue pumpable.

A much larger ASDA unit has been constructed at the H-Coal pilot plant at Catlettsburg, Ky. However, the unit has not been in operation at all. This is because the Catlettsburg plant has not been operated in the boiler fuel mode for which the ASDA unit was designed as a system for solids/liquid separation of residual slurry obtained after coal liquefaction.

2.4.1.2 LC-Fining Process

In 1975, Cities Service and Lummus opted to market a proprietary commercially proven petroleum residue upgrading process based on expanded-bed hydroprocessing and designated it as the LC-Fining process. From May, 1976 through September, 1979, the Department of Energy (DOE) sponsored the LC-Fining process development program to determine the feasibility and optimum operating conditions for upgrading coal extract to higher quality liquid products.

The experimental work was carried out in the L-C Fining hydrocracking unit located at the New Brunswick Laboratory of CE Lummus. The unit consists of two expanded-bed reactors in series with a total liquid throughput of about 1.6 L/h. The experimental tests were made using filtered SRC produced from Kentucky coals (Nos.9 & 14) at the SRC-I Fort Lewis pilot plant. The high melting point and thermal instability of the SRC necessitated the use of a heavy residual creosote oil as a solvent.

2.4.1.2 LC-Fining Process (cont'd)

Initial runs were aimed at determing a preferred concentration of solvent, process temperature, space velocity and catalyst deactivation rate, using Cyanamid 1442-B cobalt molybdate catalyst.

The conclusions from these runs were as follows:

- conversion, denitrogenation and deoxygenation decreased with increasing SRC concentration in the feed blend and space velocity
- desulfurization is satisfactory at all levels
- no apparent catalyst aging effect on conversion, nitrogen and sulfur removal were observed. Oxygen removal showed a declining rate with aging.

A survey was made of commercial catalysts to improve nitrogen removal. A modified Co/Mo (1442 B) from American Cyanamid was the reference catalyst and was tested in parallel with two Ni/Mo and one Ni/W catalyst. A goal of 0.3 wt% nitrogen was set for the fuel oil fraction (200-460°C). The results show that the Ni/Mo catalyst (Shell 324) was the most effective. The fuel oil fraction (bp 200-460°C) met the target of 0.3% N. In addition, this catalyst provided a greater conversion of heavier fraction to lighter product coupled with a higher recovery of total liquid fuel without any increase in hydrogen consumption.

2.4.1.2 LC-Fining Process (cont'd)

When SRC containing process-derived solvent was hydrogenated using Ni/Mo (Shell 324) catalyst, about 85% of the feed was converted to distillate. The reduction of heteroatoms ranged from 87% for nitrogen and up to 95% for oxygen.

2.4.1.3 Integrated Two-Stage Liquefaction

An integrated two-stage coal liquefaction process development unit (PDU) having 0.3 t/d coal feed capacity has been operating at the Lummus, New Brunswick, NJ, facilities since May 1980. The process, as it is now being investigated, is shown in Figure 2-7.

Coal, hydrogen, and recycle solvent react in the SCT reactor. The SRC-I preheater having a higher outlet temperature is used as an SCT reactor. The gases and light oil are flashed off while the heavy oil goes to the deasher. The deasher underflow, containing virtually all of the ash and the insoluble organic matter (IOM) goes to a vacuum flash, where the distillate portion goes overhead to a continuous distillation column. The ash-laden flashed bottoms are stored in drums. In a commercial plant, this stream would be gasified to generate hydrogen for the process.

The deasher overflow goes to an atmospheric flash. The overhead is combined with the vacuum flash overhead as feed to a continuous distillation column to recover the antisolvent.
The atmospheric flash bottoms constitute the feed to the LC-Finer, along with make-up hydrogen. The LC-Finer product is atmospherically flashed to collect light distillate product. The heavy oil product is then vacuum flashed to collect heavy distillate product and the bottoms are recycled as solvent. In this manner, the only portion of extract (> 460°C) leaving the

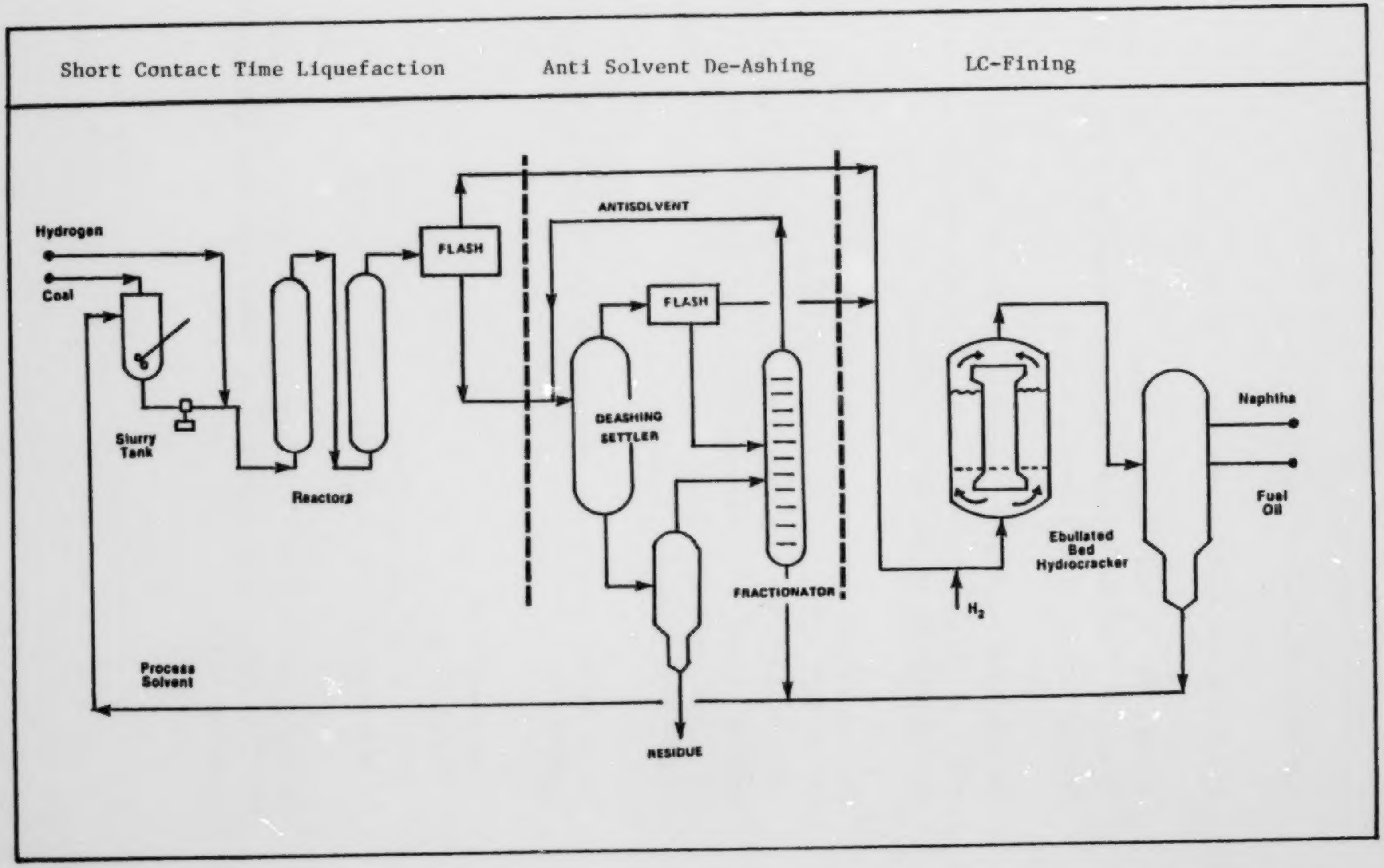

Figure 2-7: Simplified Two Stage Liquefaction Process Flow Diagram

2.4.1.3 Integrated Two-Stage Liquefaction (cont'd)

process is in the vacuum flashed bottoms from the deasher underflow. All the extract that is not converted in the LC-Finer becomes part of the SCT recycle solvent.

The PDU runs have mainly investigated solvent self-sufficiency, effect of process variables on product distribution, and anti-solvent deashing of the reactor effluent.

The SCT reactor has been operated at an outlet temperature of 457 to 463°C and an outlet pressure of about 16.5 MPa. The coal space velocity has been varied from 24 to 59 kg coal/min/m^3 of reactor volume. The start-up solvent was Koppers Creosote Oil which was gradually displaced by process derived containing 18-48 wt% of heavy oil ($>$ 460°C).

The LC-Fining reactor was operated at the following range of conditions:

Temperature, °C	410 to 432
Pressure, MPa	18.6
Hourly space velocity, V/h/V	0.2 to 0.6
Concentration of $>$ 460°C feed, vol.%	30 to 60

The short contact time SCT reactor yields along with the integrated TSL yields are presented in Table 2-8 for Indiana V bituminous coal.

Table 2-8 SCT Reactor and Integrated TSL Yields

	TSL Reactor	Integrated TSL (SCT + LC-Finer)
Feed	Coal + Solvent	
Operating Conditions		
Temperature, °C	463	
Pressure, MPa	16.5	
Solvent to Coal Ratio	1.8	
Coal feed rate, kg/s/m^3	224	
Space velocity, V/h/V	-	
Yields*, wt% MF Coal		
$H_2O + H_2S + NH_3$	4.23	12.44
CO_x	1.15	1.15
$C_1 - C_4$	2.84	5.58
C_5 - 200°C	0.25	6.88
200 - 260°C	1.82	11.75
260 - 345°C	1.02	24.97
345 - 460°C	9.85	5.33
> 460°C	61.48	18.60
Ash + IOM	17.89	17.78
Sub-total	100.53	104.44
Hydrogen Consumption	0.53	4.44
Total	100.0	100.0

* Indiana V Bituminous Coal.

Source: H.D. Schindler et al., "The Integrated Two-Stage Liquefaction Process", presented at AIChE Annual Meeting, New Orleans, La., Nov. 1981.

2.4.1.3 Integrated Two-Stage Liquefaction (cont'd)

The SCT reactor yields of light hydrocarbon gases (C_1-C_4) is about 2.8 wt% MF coal. The yield of recycle solvent (260 to 460°C fraction) is about 11 wt% whereas the hydrogen consumption is less than one percent. The low hydrogen consumption, as verified by Conoco Coal Development Corporation (through solvent and product analyses) is due to the fact that hydrogen donor capabilities of the recycle solvent provides as much or more hydrogen to the SCT hydroliquefaction reaction as does gaseous hydrogen.

The LC-Finer reactor yields are given in Table 2-9. The C_1-C_4 gas make is about 2.2 wt% of feed but it would increase with catalyst age. The results shown in Table 2-9 correspond to the catalyst age of 1310 expressed as kilograms of > 460°C feed processed per kilogram of catalyst. The yields of total liquid distillate (C_5- 460°C) is about 70 wt% of LC-Finer feed with the total hydrogen consumption of 3.7 wt%. It is evident that all of the hydrogen consumption is shifted to the second-stage operation (LC-Fining).

The PDU yields for an integrated TSL liquefaction are also shown in Table 2-8. The yield of C_1-C_4 hydrocarbon gases is about 5.6 wt% MF coal. The total liquid yield of 67.5% is achieved with total hydrogen consumption of 4.4%. About 7% of this liquid is naphtha.

2.4.2 **Overall Conceptual Plant Configuration**

The overall conceptual process configuration of a self-sufficient TSL plant is being designed by the process developer. However, the results are not be available yet.

Table 2-9 LC-Finer Reactor Yields

Feed	345 - 460°C
Operating Conditions	
Average Reaction Temperature, °C	416
Reactor Pressure, MPa	18.6
Space Velocity, V/h/V	0.37
Catalyst Life, kg Feed/kg Catalyst	1310
Yields, wt% Feed	
$H_2 + H_2S + NH_3$	6.37
C_1-C_4	2.25
C_5 - 200°C	12.57
200 - 260°C	10.35
260 - 345°C	22.70
345 - 460°C	23.94
> 460°C	25.47
Sub-total	103.67
Hydrogen Consumption	3.67
Total	100.0

Source: H.D. Schindler et al., "The Integrated Two-Stage Liquefaction Process", Presented at AIChE Annual Meeting, New Orleans, La., Nov. 1981.

2.4.2 Overall Conceptual Plant Configuration (con'd)

The process developer has provided details of only three key units of a 30,000-t/d conceptual plant for Indiana V bituminous coal. These units are: (i) Short Contact Time (SCT) hydroliquefaction, (ii) Antisolvent Deashing and, (iii) LC-Fining. A simplified block diagram, indicating the overall material balance for the three units is shown in Figure 2-8.

In the conceptual 30,000 t/d Integrated Two-Stage Liquefaction plant, dried and pulverized Indiana V coal is slurried with a recycle solvent to a 36 wt% concentration. The recycle solvent contains both distillable and non-distillable constituents and is derived from both the Antisolvent Deashing (30%) and LC-Fining (70%) units.

The coal slurry is pumped to the SCT section of the plant where make-up and recycle hydrogen are added prior to entering the fired heater-reactor. In this heater, dissolution of the coal occurs followed by thermal reaction of the solution resulting in approximately 91% coal conversion. Heater outlet temperature is maintained at approximately 460°C. The main product from the SCT heater-reactor is a non-distillable extract which is upgraded in the L-C Fining unit, after solid/liquid separation in the ASDA unit. Net hydrogen consumption in the SCT unit is 1.3 wt% MF coal and in the subsequent LC-Fining unit is 1.7 wt% of extract-feed, which is equivalent to 3.0 wt% MF coal.

2.4.3 Technical Review

2.4.3.1 Types of Coal Investigated

The coal tested so far has been Indiana V bituminous coal. This coal contains 2.6 wt% moisture and 10.4 wt% ash on "as-received"

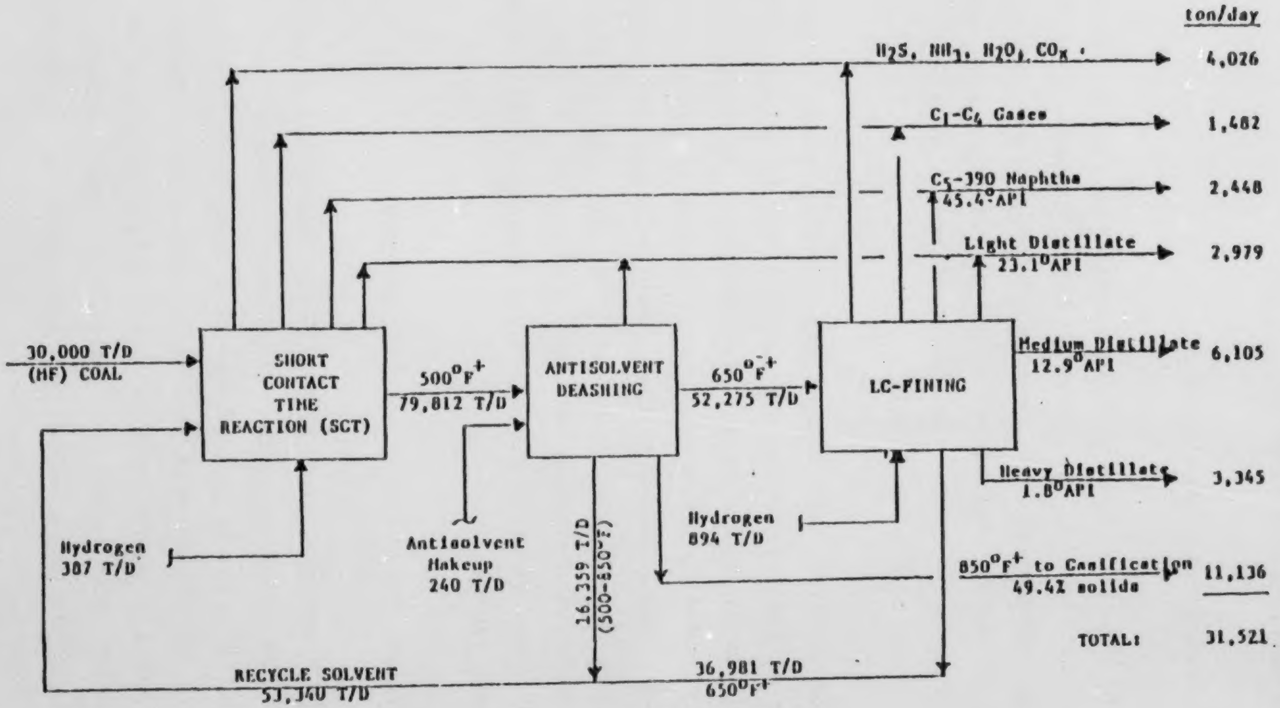

Figure 2-8: Overall Material Balance Diagram of 30,000 t/d Integrated Two Stage Liquefaction Plant

Source: C.E. Lummus Company, Bloomfield, N.J.

2.4.3.1 Types of Coal Investigated (cont'd)

basis (see Appendix A for complete analysis). The second coal scheduled for testing is Illinois No. 6 bituminous coal. No laboratory experiments have been made either with sub-bituminous or lignitic coals. According to the negotiations with DOE, the process developer's plans include experimental runs with Wyodak sub-bituminous coal during 1982.

It may be pointed out that short contact time liquefaction behaviour of coals of different ranks cannot be predicted in the absence of experimental data. The consensus is that intrinsic reactivity actually decreases with increasing rank over the entire rank range. However, the common criterion for reactivity is the production of soluble material and low rank coals require a good source of hydrogen, so some workers (researchers) have concluded that these coals are less reactive than those of intermediate rank.

2.4.3.2 Product Yields, Distribution and Efficiencies

Product yield and distribution data for a conceptual plant operating with Indiana V bituminous coal were presented earlier in Figure 2-8. It must be pointed out that results in Figure 2-8 do not reflect the yields of a self-sufficient plant. These yields represent gross products from the plant before allowance for fuel and utilities requirements. Based on these results about 3.4 bbl/t liquid (C_5- 460°C) on MAF basis is produced from Indiana V Coal in the TSL process

The product yields for Saskatchewan lignits cannot be estimated because lignites have not been processed in the TSL process, and sufficient data for prediction are not available.

2.4.3.3 Product Quality

The product streams from the TSL process have not been fully characterized nor their quality has been ascertained. However, the specific gravity and the elemental analysis has been determined and is presented in Table 2-10.

From the elemental analysis standpoint, the TSL liquids contain very low sulfur (0.05 to 0.09 wt%), moderate nitrogen (0.1 to 0.2 wt%) and oxygen contents (0.3 to 0.7 wt%).

2.4.3.4 Status of Research and Development

The TSL process was partly demonstrated at the SRC-I Wilsonville pilot plant. A demonstration plant for deashing of residual slurry from the 600-t/d H-Coal pilot plant (in boiler fuel mode) was erected at Catlettsburg, KY. The H-Coal pilot plant has not been operated under boiler fuel mode due to cut backs in their program. Therefore the TSL pilot plant was not operated.

A two-stage liquefaction pilot plant, 0.5 t/d coal, consisting of short contact time (6 minutes) liquefaction reactor, ASDA unit and LC-Fining ebullated-bed hydrocracker is in continuous operation at the New-Brunswick, N.J. facilities of CE-Lummus Company. Extensive experiments to process various bituminous and subbituminous coals and to evaluate catalyst life as well as other process variables studies are being carried out. Currently the funding for operation of this plant has been provided by the U.S. Department of Energy (DOE) till the end of 1982. However, negotiations with DOE are in progress for the extension of funds till 1984. The process developer has yet to carry out techno-economic evaluation of a conceptual, integrated Two-Stage Liquefaction (TSL) plant.

Table 2-10 Specific Gravity and Elemental Analysis of TSL Product Streams

	Naphtha C_5 -200°C	Light Distillate 200 - 260°C	Mid-Distillate 260 - 345°C	Heavy Distillate 345 - 460°C
Sp. Gravity, °API	45.4	23.1	12.9	1.8
Elemental Analysis, wt%				
Carbon	86.04	87.87	89.30	90.77
Hydrogen	13.17	11.26	10.26	8.47
Oxygen	0.59	0.68	0.27	0.45
Nitrogen	0.11	0.12	0.12	0.23
Sulfur	0.09	0.07	0.05	0.08

Note: Products obtained from 0.3 t/d PDU operating with Indiana V bituminous coal.

Source: H.D. Schindler et al., "The Integrated Two-Stage Liquefaction Process", resented at AIChE Annual Meeting, New Orleans, La., Nov. 1981.

2.5 SOLVENT REFINED COAL PROCESS

The Solvent Refined Coal (SRC) process converts coal into distillate hydrocarbons (C_5-460°C) as main products. Low-sulfur, ash-free solid SRC and C_3/C_4 LPG are the co-products. The main features of the SRC-1 process are:

- Crushed coal is liquefied in a non-catalytic tubular reactor in the presence of hydrogen gas and the process-derived solvent.
- The reactor is operated at 440°C, and 11.4 MPa with a residence time in the order of 30 to 60 minutes.
- Reactor effluents are flashed to yield a gaseous stream and a liquid-solid stream containing undissloved coal and ash.
- The liquid/solids stream is deashed using a proprietary Kerr-McGee Critical Solvent Deashing (CSD) process.
- The deashed liquid (which is molten SRC) is catalytically hydrocracked using LC-Fining technology developed jointly by Lummus and Cities Service Company.
- The residue from Kerr-McGee CSD unit is gasified to produce make-up hydrogen.

2.5.1 **History and Development**

The Solvent Refined Coal (SRC) process, which actually originated in Germany in the early 1900's, is a mild hydrogenation coal extraction process. In the early 1960's Pittsburg & Midway Coal Mining Company (P&M), a subsidiary of Gulf Oil corporation, began the development of the SRC process and intended to produce a

2.5.1 **History and Development** (cont'd)

low-sulfur deashed solid fuel as a power plants feed. Hydrogenation was, therefore, held at a minimum required for sulfur removal only, with the amount of liquids produced being incidental to the operation. In the early 1970's two SRC pilot plants were built to carry out further process development: a 5-tonne-per-day plant located at Wilsonville, Alabama and a 45-tonne-per-day unit located at Fort Lewis, Washington. Air Products subsidiary of Catalytic Inc., engineered, built, and operated the 5-t/d plant while Wheelabrator-Frye's subsidiary, Rust Engineering, designed and built the 45-t/d unit. Both of these plants were built under the funding from the U.S. Dept. of Energy (DOE). In 1975, laboratory studies carried out by P&M revealed that coal slurried in a mixture of process solvent, SRC, and ash could be dissolved, desulfurized, and hydrocracked to hydrocarbon gases and distillate fuels under severe hydroprocessing conditions in a single-stage reactor so-called dissolver. Subsequently, the Fort Lewis pilot plant was modified to permit slurry recycle in order to convert coal into liquids. Thus, the original SRC process was renamed as SRC-I (solid fuel mode) and as SRC-II (liquid fuel) processes.

The SRC-I process has undergone a number of modifications. The initial version, which incorporated mechanical filtration for deashing the product slurry, has been replaced with a non-mechanical separation systems such as Lummus' "Antisolvent Deashing" and Kerr McGee' "Critical Solvent Deashing" technologies. Another significant change, or rather extension of the process, is an addition of catalytic hydroprocessing step, to produce distillate product from solid SRC product. The process is, therefore, often referred to as "two-stage SRC-I process".

2.5.1 **History and Development** (cont'd)

In January 1977, the Wilsonville pilot plant was modified to accommodate a Kerr-McGee Critical Solvent Deashing (CSD) unit for solids-liquid separation of vacuum flash bottoms. Figure 2-9 is a flow diagram of the CSD unit. The decision was based on bench-scale investigation (0.7 kg/h) at Kerr-McGee headquarters which indicated that continuous deashing of SRC vacuum bottoms was a more promising approach than batch filtration.

Critical solvent deashing employs three unique characteristics of critical solvents to effect solids-liquid separation, specifically:

- Density of the critical solvent changes rapidly with temperature.
- Dissolving power of critical solvent changes approximately in proportional to its density.
- Dissolving power of a critical solvent is greatest for "lighter" coal fractions and decreases for "heavier" coal fractions.

The C-E Lummus antisolvent deashing (ASDA) technology was also investigated for SRC-I process. Testing was carried out at Fort Lewis pilot plant for four months.

Comparative performance results of Kerr-McGee CSD and CE-Lummus ASDA units are given in Table 2-11. The Kerr-McGee CSD process was selected. The reason cited was that "the Lummus process is

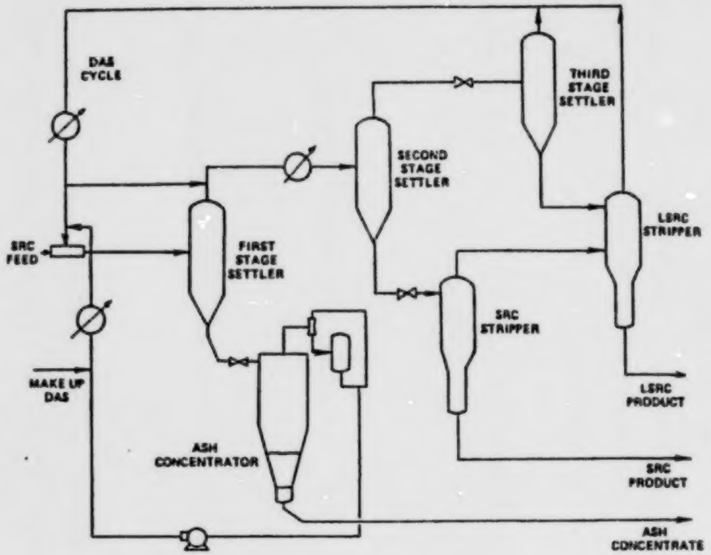

Figure 2-9: Kerr-McGee Critical Solvent De-Ashing

Source: R.M. Adams et al., "Critical Solvent Deashing of Liquefied Coal", Chemical Engineering Progress, pp. 44-48, June 1979

Table 2-11 Kerr-McGee Critical Solvent and Lummus Antisolvent Deashing Performance Results in SRC-I Process

	Kerr-McGee CSD	Lummus ASD
Feed source	Vacuum Tower Bottoms	758-1034 kPa Stripper Bottoms
Feed composition, %		
SRC	73.0	62.5
Ash	12.0	5.0
Unconverted coal	8.0	2.5
SRC Process Solvent	7.0	30.0
Ash in the SRC Product, %	0.1	0.16
SRC Product Recovery, %	81.7*	75.0**
Ash Residue Composition, %		
SRC	34.5	48.0
Ash	36.0	33.3
Unconverted coal	24.0	16.7
SRC Process Solvent	4.0	2.0
Deashing Process Solvent	1.5	2.0

* Considering solvents left in the residue as an additional SRC loss.

** Considering all pyridine soluble material left in the ash residue as SRC loss.

Source: H.M. Risbud, "Technical Comparison Among Solids Separation Alternatives in SRC-I Process", EPRI Contractor's Conference, Palo Alto, CA, May 14, 1981.

at an earlier stage of process evolvement than the Kerr-McGee process. Lummus has not yet demonstrated a sustained operation that produces an acceptably low ash in the overflow and a high ash in the underflow and a high enough (70% plus) SRC recovery".

2.5.2 SRC-I Demonstration Plant

On July 1978, U.S. DOE contracted with Southern Company Services for a four-phase program to demonstrate the SRC-I process technology (new version) so that it can be commercialized. The principal subcontractor is International Coal Refining Company (ICRC) which is a joint venture company of Air Products and Chemicals Inc. and Wheelabrator Frye, Inc. The first unit of the SRC-I Demonstration Plant, to be built near Newman, Kentucky, will be designed to process 5445-t/d (6000 short tons/day) of coal.

A simplified process flow diagram of the 5445-tonne-per-day SRC-I demonstration plant is presented in Figure 2-10. The coal preparation section (not shown) includes receiving, unloading, conveying, storing, reclaiming, drying, and grinding the coal as feed for the process units. Approximately 86% of the coal is ground, dried, and fed to the SRC Process Unit, and the remaining 14% after grinding and drying is fed to the Gasification Unit.

In the SRC process unit, the coal is slurried in a process solvent, pumped to reaction pressure, mixed with a hot, hydrogen-rich gas stream, and heated, first against hot returning process solvent, and then in a fired-heater. Within the fired-heater, coal dissolution is accomplished and hydrogenation reactions begin. Additional hydrogen-rich gas is added at the

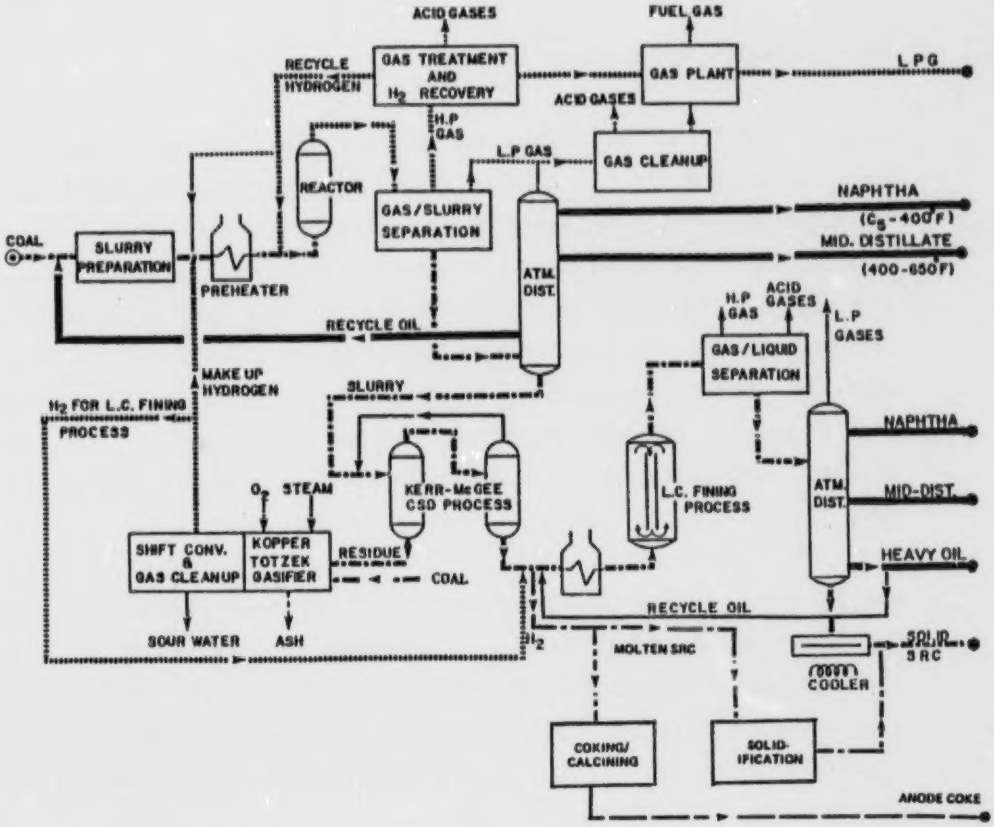

Figure 2-10: Simplified Process Configuration of 5445 t/d SRC-I Demonstration Plant

2.5.2 **SRC-I Demonstration Plant** (cont'd)

exit of the fired heater and the mixture flows to the dissolvers, where hydrogenation and desulfurization reactions are completed. The high-pressure hydrogen-rich gas is separated from the product slurry, which is flashed and distilled to remove process solvent and lighter components. The remaining SRC/ash slurry is sent to the Kerr-McGee Critical Solvent Deashing unit, where it is mixed with the deashing solvent and separated into a molten SRC stream and a solids residue stream. The Kerr-McGee solids residue stream, containing ash, undissolved coal, and about 20% of the SRC, is sent to the coal gasification unit.

Gasification serves two purposes: (1) it converts the residue into an inert slag; and (2) through gasification of the carbonaceous components of the ash concentrate plus some supplemental coal, it generates the make-up hydrogen that is required for both the liquefaction and the expanded-bed hydrocracking (LC-Fining) processes.

The resulting gas is compressed and passed over a shift conversion catalyst to produce the raw make-up hydrogen stream. After the acid gases are removed, the purified hydrogen stream is divided, with approximately two-thirds going back to the SRC Process Unit and one-third going to expanded-bed hydrotreating (LC-Fining process).

Following its use in each of these units, the high-pressure hydrogen-rich gas passes though a series of treating steps before it is recycled to the SRC Process Unit. The combined-gas stream is washed to remove chlorides and ammonia, scrubbed to remove

2.5.2 **SRC-I Demonstration Plant** (cont'd)

acid gases, dried, and cryogenically processed to recover light hydrocarbons, which are used as plant fuel. Acid gases recovered after treating the combined-gas stream and the make-up hydrogen stream are sent to the Sulfur Recovery Area, where molten sulfur is produced.

According to the present plans, the molten SRC stream will be processed in three separate units. The expanded-bed hydrocracking process would receive between one-third and two-thirds of the molten SRC and convert it catalytically into naphtha, fuel oil, and low-sulfur solid SRC. Another third of the molten SRC would go to coking and calcining to produce anode coke for aluminum manufacture, and the balance of the molten SRC will be sent to the solidification unit to produce solid SRC.

2.5.3 **Overall Conceptual Plant Configuration**

The overall process configuration of the commercial plant will be the same as that of the 5445-t/d SRC Demonstration Plant. Desired capacity of the commercial plant will be attained by adding the required number of trains to the SRC demonstration plant. However, in the commercial plant configuration, gasification of ash residue is planned to be carried out using Shell-Koppers technology assuming that this technology will be advanced to commercial readiness by 1985.

2.5 SOLVENT REFINED COAL PROCESS (cont'd)

2.5.4 **Technical Review**

2.5.4.1 Types of Coal Investigated

Different types of bituminous coal have been processed in bench-scale and PDU units at Harmarville and in the 45-t/d Fort Lewis pilot plant under SRC-II process configuration. Exploratory bench-scale tests with Wyodak subbituminous coal were carried out at Merriam laboratory in Kansas. Elemental and proximate analyses of these coals are given in Appendix A.

In the SRC-I pilot plant, extensive testing has been done with the Kentucky No. 9 and No. 14 mixture, Pittsburgh No. 8, and Illinois No. 6 coals. Amongst these, Illinois No. 6 coal has been found to be most reactive.

The lignitic coals have not been tested. The laboratory-stage testing of Wyodak coal indicated that addition of pyrites is required because this coal contains only 1.1% sulfur. The sulfur content of Saskatchewan lignites is even lower (0.9% in Willow Bunch and 0.6% in Estevan on MF basis).

2.5.4.2 Product Yield, Distribution and Efficiencies

The liquefaction results obtained from processing of various types of coal in SRC-I pilot plants indicate that the total liquid yields from bituminous coal with high-sulfur content (up to 5.2%) are higher when compared with those obtained from low sulfur (1.1%) Wyodak subbituminous coal. Laboratory studies, however, have also demonstrated that liquid yields from low-sulfur coals can be improved by the addition of pyrites.

2.5.4.2 Product Yield, Distribution and Efficiencies (cont'd)

Typical reactor yields (see Table 2-13) from Kentucky 9 bituminous coal indicate about 15.0 wt % MF coal of total liquids along with 58 wt % MF coal of solid SRC is produced. If the solid SRC is further hydrocracked, the total liquid yields would increase to 64.3 wt % MF coal.

The product yields and distribution for a conceptual self-sufficient commercial plant operating with Kentucky 9 coal are presented in Table 2-14. The basis used is 25,000 t/d of "as received" coal. The hydrotreating of SRC is carried out at high severity (85% conversion of SRC to liquid products). Approximately two-thirds of the liquid products are produced in the hydrocracker with the remaining one-third being produced in the dissolver.

The overall and liquid products efficiencies are 65.6% and 53.7% respectively.

The utility requirements from outside source are water and electricity. About 220 MW of electricity is purchased from outside. The raw water requirements are estimated between 315 to 440 L/s.

2.5.4.3 Product Quality, Utilization and Upgrading

The liquid products are naphtha (C_5-200°C), mid-distillate (200-345°C) and heavy-distillate (345-460°C). About 37% of total naphtha, 29% of total mid-distillate and 25% of total heavy distillate are produced in the dissolver. The balance is produced in LC-Fining.

Table 2-13 Typical Reactor Yields in the SRC-I Process

	Fort Lewis Pilot Plant	Demonstration Plant Design Basis
Operating Conditions		
Pressure, MPa	10.7	11.7
Temperature, °C	454	440
Solvent to Coal Ratio	1.66	1.6
Reactor space velocity w/h/v	53	50
Yields, Wt% MF coal *		
CO_x	1.3	1.0
H_2S	2.1	1.9
$H_2O + NH_3$	4.7	4.3
C_1	2.2	2.7
C_2	1.3	1.8
C_3	1.1	1.5
C_4	0.8	0.4
C_5-200°C	6.6	4.6
200-345°C	8.1	4.7
345-460°C	8.1	6.3
SRC	58.3	56.5
Unconverted coal	6.6	6.7
Ash	9.3	10.4
Sub-total	102.4	102.8
Hydrogen Consumption	2.4	2.8
Total	100.0	100.0

* Kentucky No. 9 bituminous coal

Table 2-14 Estimated Product Yield and Efficiencies in the SRC-I Conceptual Commercial Plant (Two-Step Process Alternative) [Basis: 25,000 t/d As-Received Coal]

	Kentucky No.9
Coal (As-Received)	
Moisture, wt%	10.0
Ash, wt% of MF Coal	10.4
Input	
MF Coal to Reactor, t/d	19408
MF Coal to Offsites, t/d	3052
Electricity, MW	220
Output, t/d (bbl/d)	
LPG	1555
Naphtha (C_5-200°C)	4760 (35700)
Mid-Distillate (200-345°C)	4954 (30269)
Heavy-Distillate (345-460°C)	1173 (704)
SRC Solid	1147
Sulfur	N.A.
Phenols	N.A.
Ammonia	N.A.
Liquid Product Yield (C_5-460°C)	
bbl/t As-Received Coal to Reactor	2.67
bbl/t MF Coal to Reactor	3.43
bbl/t MAF Coal to Reactor	3.83
Naphtha in Total Liquid, wt%	43.7
Thermal Efficiency, %	
Overall	65.6
Liquid Products	53.7

Note: Based on high-severity hydrocracking (LC-Fining) of SRC solid to maximize liquid yields . Naphtha and fuel oil are assumed to be 7.5 and 6.1 bbl/tonne.

Source: "Demonstration and Commercialization of the SRC-I Technology", 7th Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity, U. of Pittsburgh, Pittsburgh, Aug. 5-7, 1980.

2.5.4.3 Product Quality, Utilization and Upgrading (cont'd)

The sulfur, nitrogen and oxygen content of the liquids produced is given in Table 2-15. As can be seen, the oxygen and nitrogen contents in any of SRC liquids are high.

- Raw Naphtha

  To produce feedstock suitable for naphtha reforming into high-octane gasoline feedstock, the heteroatoms content in the raw naphtha has to be reduced by hydrotreating to about 1 ppm in order to prevent deactivation of reforming catalyst. Mild catalytic reforming tests of hydrotreated SRC naphtha have shown that such a reforming would result in high-octane gasoline blendstock, typically with octane number of 103.

The market potential for SRC-I Liquids is also being evaluated in the following areas:

- Using SRC-I naphtha as a catalytic reforming feedstock to produce benzene, toluene, and xylene for the petrochemical industry.

- Using the middle distillate (260-345°C fraction) as a hydrocracker feedstock to produce more gasoline, diesel fuel, and turbine fuel.

- Using the heavy oil (345-460°C fraction) as a feedstock for the production of carbon black.

Table 2-15 Elemental Analysis of Products from SRC-I Process*

	Naphtha (C_5-200°C)	Mid-Distillate (200-345°C)	Heavy Oil (345-460°C)	SRC Solid (> 460°C)
Sulfur	0.15	0.13	0.15	0.15
Nitrogen	0.33	0.7	0.75	1.33
Oxygen, wt%	2 - 4	2 - 4	2 - 4	0.9

Notes: * Based on 88% conversion of SRC feed in the LC-Fining Process. Values shown in parenthesis are for petroleum-derived products

Sources: (1) J. Caspers, et al, "Upgrading Solvent Refined Coal by Hydrogenation", 88th National AIChE Meeting, Philadelphia, June 8-12, 1980.
(2) S.M. Morris and E.P. Foster, "The SRC-I Coal Refining Demonstration Plant", EPRI Conference on Synthetic Fuels-Status & Directions, San Francisco, CA., October 13-16, 1980.

2.5.4.3 Product Quality, Utilization and Upgrading (cont'd)

The SRC-solid product has very low sulfur content. In addition, its low inorganic ash content (0.2%), low melting point (149°C), and a high Hardgrove grindability index (176) would permit its use in oil-designed boilers. Firing of the finely pulverised (90%$^+$, 325 mesh) SRC solid in an oil-designed boiler is now being tested. Results are not yet available.

2.5.4.4 Status and Commercialization Prospects

Since the inception of the International Coal Refining Company (ICRC), the SRC-I process development work is progressing in four phases.

During Phase 0, ICRC prepared the conceptual design of a 27,000-t/d commercial plant, preliminary cost estimates, marketing assessments, economic evaluation, and environmental appraisal. Process options were evaluated, critical technology areas requiring additional data were identified, and the economics were assessed for both a five-module "grass-roots" and a commercial plant of the same size expanded from the 5445 t/d demonstration plant. The Phase 0 work was completed in July 1979, and in October 1979, the DOE authorized Phase 1, the detailed engineering of 5445 t/d SRC-1 demonstration plant.

On 7 August 1980, a cost-sharing agreement was signed between ICRC and DOE in Owensboro, Ky, covering the remainder of the demonstration plant program through start-up and operation of the facility. Under the terms of the agreement, ICRC will invest $90 million in the project, the Commonwealth of Kentucky will

2.5.4.5 Status and Commercialization Prospects (cont'd)

invest $30 million, and the DOE will fund the balance. The contract states that ICRC will eventualy own the coal refinery after buying out the federal and state governments' interest. At that time, ICRC plans to expand the facility fivefold to 27,000 t/d.

Under the cost sharing agreement, ICRC became the prime contractor on the project. As an important subcontractor to ICRC, Southern Company Services will continue its pioneering role in the development of the SRC technology by providing broad technical reviews and product use studies.

Since the signing of the cost-sharing agreement, Alcoa and Cities Services Co. have agreed in principle to become minority partners of ICRC.

At present, essentially all major technology selections have been completed, and the engineering is moving rapidly. Also, the Construction Manager/Constructor selection is imminent. ICRC has five engineering subcontractors: Air Products, Catalytic, CE-Lummus, R.M. Parsons, and Rust Engineering. Site work started in early 1981. The procurement and construction (Phase II) was also started in 1981. The operation and evaluation of the plant (Phase III), according to the present schedule, is expected to commence in 1986.

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2.6 ZINC-HALIDE PROCESS

The Zinc-Halide catalytic process is designed to convert bituminous and subbituminous coal into distillates (in the gasoline range) by severe catalytic hydrocracking. The process may be applied either to coal in a one-step process or to hydrocracking of coal extract in the second step of a two-step process. The Zinc-Halide process has the following main features:

- Coal is hydrocracked at 413°C and 13.8 MPa (2000 psig) in the molten pool of zinc chloride.
- Hydrocarbon liquids are distilled, and material boiling above 200°C is returned to the process for reaction to extinction.
- The reactor underflow containing unconverted coal, ash, and heteroatom compounds (e.g., ZnS, ZnO), as well as spent catalyst is sent to a catalyst regenerator.
- Catalyst-regenerator is a fluidized bed combustor where HCl is added to the feed air to convert ZnO into $ZnCl_2$.
- The regenerator is operated at 930°C and about 103 to 483 kPa (15 to 70 psig) pressure to vaporize zinc chloride.
- Ash is removed from the zinc chloride vapors using cyclones and the condensed $ZnCl_2$ is recycled back to hydrocracker.

2.6.1 **History and Development**

The early use of metal halides for coal hydrogenation has been documented by H.H. Storch. This involved a small amount of metal halide (<5%) impregnated on coal, and was tested in pilot and commercial plants in Germany, Great Britain, Japan and the United States. Under these conditions halogen-promoted tin catalysts were the most active. The use of large quantity of zinc chloride for coal liquefaction was developed at the Consolidation Coal Company which was looking for a more active catalyst than the hydrofining type (Co, Mo, Ni and Al_2O_3) catalysts employed to upgrade heavy oils and coal extracts. Zinc chloride, zinc bromide, aluminum bromide and stannous chloride were tested as molten salts in large amounts.

Zinc chloride was found relatively more active than other catalysts and therefore it was selected for further investigation under the Consolidation Coal Company contract, with the Office of Coal Research (OCR). During 1963 to 1968 OCR contract period, a continuous zinc chloride hydrocracker was built and operated on deashed coal extract. Results of this work led to a cost-shared program carried out by DOE, Conoco, and Shell Development Company. This contract ran from 1975 to 1980. It involved first demonstration of the process concept on a bench-scale level. It was shown in particular that a coal slurry could be fed directly to a molten zinc chloride reactor and that a regeneration of the spent catalyst could be achieved. The design and engineering of a 1-tonne-per-day process development unit (PDU) began in March 1975 and the construction was completed in May 1978. The PDU, which is located at the Conoco Research Center in Library, Pennsylvania, was operated until late 1979. At this time the major work on the project was terminated.

2.6.1 **History and Development** (cont'd)

Bench-scale investigation was carried out in a 1-kg/h of coal continuous hydrocracker with a $ZnCl_2$ catalyst. The main objectives was to conduct process optimization study. The variables and the ranges at which they were investigated are:

Temperature, °C	385, 400, 413, 427, 440
Pressure, MPa	16.5, 20.7, 24.1, 31.0
Residence time, min.	30 to 190
$ZnCl_2$ to MF coal ratio	0.8, 1.0, 1.8
H_2 feed rate, m^3/kg coal	1.8 to 2.0

The hydrocracker feed consisted of pulverized coal (147 μm) slurried in 92 wt% benzene or toluene and 8 wt% polystyrene. The solvent refined coal (SRC) received from the SRC-I pilot plant at Ft. Lewis, Washington was also used as feed for several runs. Few runs were carried out using a two-stage hydrocracker reactor (connected in series). After each run, the boiling point distribution of the 475°C distillate and the MEKS (methyl ethyl ketone-solubles) were measured. The higher molecular weight products (in general) have higher boiling point. Therefore, boiling point distribution curve was used to examine the effect of reaction conditions upon the product from hydrocracking of coal. The following major conclusions were drawn from the process variables study:

- Increasing conversion at a given temperature by increasing residence time decreases the average molecular weight of the composite > 480°C product.

2.6 ZINC-HALIDE PROCESS (cont'd)

2.6.1 **History and Development** (cont'd)

- Increasing conversion by increasing the hydrocracking temperature progressively decreases the average molecular weight of composite > 480°C product.

- The data obtained at 16.6 MPa and 23.2 MPa were comparable. The molecular weight distribution of the composite 480°C product observed at the lower pressure was found comparable to that observed at the higher pressure, indicating that hydrogen partial pressure does not affect the molecular weight of the liquid products produced.

- The average molecular weight of the product from two-stage hydrocracking was lower than that from single stage. This result suggests that in multi-stage commercial operation an increased yield of lower boiling distillate fuel may be observed from that predicted from the "single-stage" results.

Four metal chlorides (NaCl, KCl, $CaCl_2$ and $FeCl_2$) were separately added to zinc chloride in the weight ratio of 0.2 (17 wt % of metal chloride in the salt mixture) and the reactions were carried out at relatively mild conditions 400°C temperature and 13.8 MPa pressure. These runs were made because alkali metals such as Na_2O, K_2O present in the coal ash are believed to reduce the activity of zinc chloride catalyst. These results indicate that the coal conversion (expressed as wt% MAF feed) drops from 68.9% to 31.5% when $ZnCl_2$ is contaminated with NaCl. The contamination with other salts such as KCl or $FeCl_2$ reduces the conversion to about 38%.

2.6.1 **History and Development** (cont'd)

A simplified block diagram of the 1-t/d process development unit (PDU) is shown in Figure 2-11. The unit consisted of five major blocks, namely; (i) feed preparation section, (ii) hydrocracking, (iii) catalyst regeneration, (iv) hydro let-down and, (v) product fractionation.

In the feed preparation section, coal is crushed to 147 μ m and dried in an existing Williams roller mill system to about 3% moisture. Pulverized and dried coal (or coal extract) is then mixed with solvent in an agitated tank. The feed rate can be varied from 23 to 68 kg/h depending upon coal to solvent ratio. Zinc melt is charged [illegible] rates varing from 45 to 136 kg/h.

In the hydrocracking block, two reactors were provided. They were designed to be used either separately or in series. The first reactor is an agitated vessel utilizing eight 12.7-cm diameter turbine impellers mounted on a common shaft and separated by cross-sectional baffling. The stirred reactor has 18.4 cm and 366 cm height with each stage being 38.1 cm high.

The second reactor is a gas-agitated vessel, is 12.7 cm, 762 cm height. The recycle gas rate (61,000 SCFH or 1727 m^3/h) gives a maximum superficial gas velocity of 0.43 metres/s at 24.1 MPa and 427 °C.

The hydrocrackers operate at about 427 °C. All fluids are preheated before being charged to hydrocrackers even though the reaction is exothermic. This is necessary because the reaction initiates only when the temperature reaches about 370 °C. In

2.6.1 **History and Development** (cont'd)

contrast to the bench-scale unit, two new process steps were added to the hydrocracking section: The absorber was provided to minimize the loss of HCl gas from hydrocracking, and the zinc melt still system operated under vacuum to flash off excess hydrocarbons from the spent melt. Zinc melt let-down is accomplished in two stages to minimized pressure drop across the control valve.

The catalyst regeneration section is designed to be fed by a nominal 59 kg/h of spent melt to the section with a maximum feed rate of 118 kg/h. The scale-up from the bench unit is about 40 to 1 based on both nominal feed and reactor volume. The fluidized-bed area is 23 cm, and 122 cm height with an expanded solids disengaging area over the fluidized-bed.

The refractory-lined catalyst regenerator reactor is operated at 1040°C. Operation of this reactor is similar to that of the bench-scale unit. Spent and recycle melt is fed into a deep fluidized-sand-bed. Combustion air and HCl fluidize the bed and provide a relatively high HCl partial pressure to convert spent melt back to zinc chloride. The regenerator effluent gases pass through a cyclone dust collector to remove entrained solids.

The primary regeneration reaction is as follows:

$$ZnS + 2\ HCl + 3/2\ O_2 \longrightarrow ZnCl_2 + SO_2 + H_2O$$

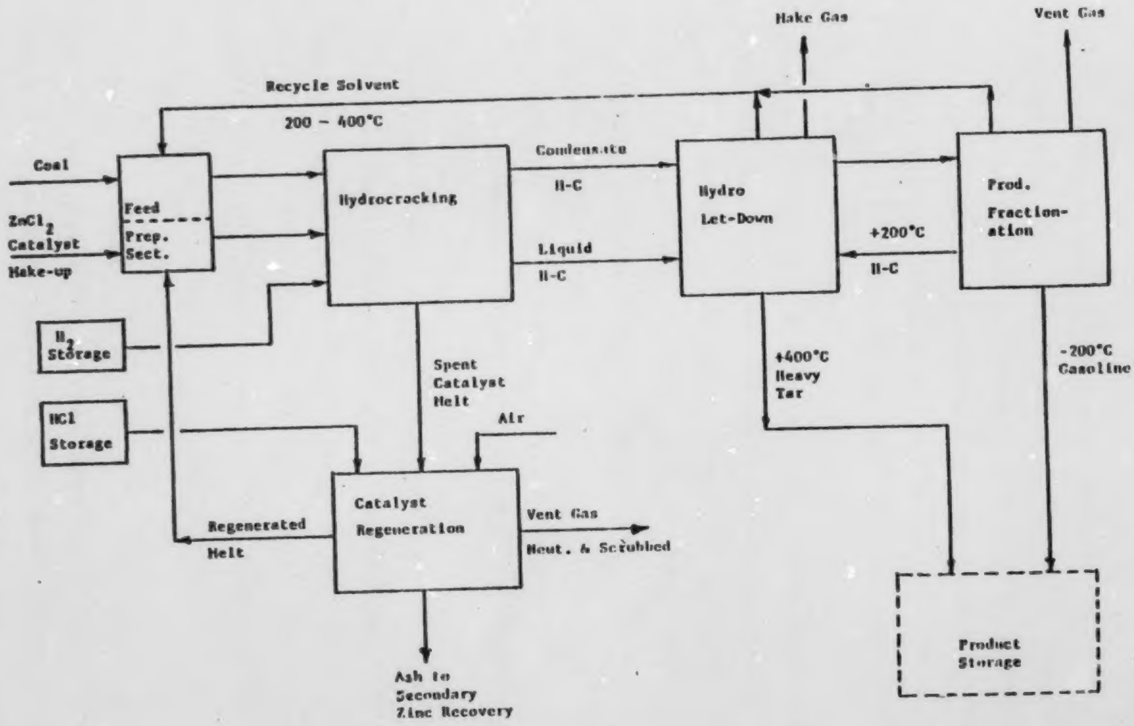

Figure 2-11: Block Flow Diagram of 1-tonne-per-day Zinc-Halide Process Development Unit

Source: DOE Report FE-1743-49, Feb. 1978

2.6.1 **History and Development** (cont'd)

The vaporized zinc chloride melt from the cyclone is recovered in a condenser. A secondary consideration in the zinc chloride melt condenser is that the liquid melt freezes in the range of 288 to 316°C. The cooling medium used is Dowtherm A vaporizing at 316°C on the shell side of the condenser. Facilities are provided for scrubbing and neutralization of flue gas before it is vented.

In the hydro let-down block, the reactor effluent gas is split into a partial condensate phase from the absorber and the final condensate phase from the main condensing train. Each of these liquid phases is let down in two steps. The high pressure let-down has two control valves in parallel with an additional bypass valve similar to melt let down. The low pressure let down has a single control valve with bypass. Hydro let-down also has a flash still to remove 480°C pitch from recycle solvent.

Recovery of products is accomplished in two continuous fractionating towers as opposed to batch-type distillation in the bench-scale unit. The first tower recovers +200°C solvent as underflow. The second tower recovers 150 to 200°C underflow solvent for recycle and overhead condensate as stabilized gasoline.

PDU Hydrocracking Runs: The PDU has been operated using a petroleum-derived aromatic solvent (Neville Solvent, bp 204°C) and SRC-1 (coal extract) made from Kentucky Coal at the Fort Lewis pilot plant.

2.6.1 **History and Development** (cont'd)

The second run was operated at 425°C which caused a high C_1-C_3 gas make of 20 wt%. However, the liquid product was over 70 wt% C_5-200°C naphtha and the carbon content of the melt was 1.9 wt% versus 5 to 6 wt% required for adiabatic operation of the regeneration. The run therefore indicated that low residual carbon levels may be obtained in the melt if a final high-temperature stage is employed.

Due to leakage problems at the reactor stirrer seal, the third run was made at lower pressure (17.2 MPa). The C_1-C_3 gas make was about 6 wt% and C_5 - 200°C liquid yield was 48 wt%.

The fourth run was operated at design capacity to evaluate the yield structure and reaction associated with recycle of process-derived > 200°C solvent. SRC and process solvent were fed for a total of 88 hours to largely equilibrate the recycle stream. The yield data presented in Table 2-16.

Catalyst Regeneration Run: The regeneration section of the PDU was operated only once due to the contract time limitation. The run indicated that the removal of impurities (C, N, S, ZnO) was between 91 - 99 wt%. In this run, the spent melt did not contain sufficient carbon to provide all the necessary heat. A propane lance was therefore added to inject propane which caused problems in uneven fluidization, combustion, and production of ZnO. The initial silica sand bed showed a high attrition rate and was later replaced by alumina. During the break-in runs, deposits formed in the overhead regenerator system; i.e., the regenerator outlet piping, the cyclone and the condenser inlet.

2.6 ZINC-HALIDE PROCESS (cont'd)

2.6.1 **History and Development** (cont'd)

The zinc oxide deposits were formed because of lower than normal HCl concentrations, and higher than normal H_2O production stemming from the use of propane. The source of the non-zinc metals is not yet understood, but appears to have come largely from being present in the feed melt.

2.6.2 **Overall Conceptual Plant Configuration**

A simplified block flow diagram for the Zinc-Halide conceptual commercial plant is shown in Figure 2-12. In the coal preparation area (not shown in this figure), feed coal is crushed and dried to 3 wt% moisture. It is then sized to 246 μm and mixed with coal-derived solvent (230 - 425°C boiling range) in the ratio of 2 parts solvent to 1 part coal. The resultant slurry is pumped to a hydroextraction reactor where it is contacted with hydrogen at a pressure of 13.8 MPa (2000 psig) and in the temperature range of 427 to 454°C for a sufficient period of time to obtain the desired conversion of coal to extract.

In the residue separation block, gases and distillate products from the hydroextraction reactor are separated from the extract, residue, and solvent via a pressure let-down system and separators. The resultant extract-residue-solvent slurry is sent to a two-stage hydroclone system consisting of 76 mm hydroclones. Here, the extract and extraction residue are separated.

Table 2-16 PDU Versus Bench-Scale Results of Zinc Halide Process

Feed	SRC-I Extract			Colstrip Subbituminous Coal
	Run PDU-4	PDU -4 with > 200°C recycled to extinction	Bench-Scale	Bench-Scale
Operating Conditions				
Temperature, °C	410 - 435	410 - 435	413	413
Pressure, MPa	17.2	17.2	24.1	16.5
Residence Time, min.	120	-	60	150
Product Yields, wt% MAF Feed				
$CO + CO_2 + H_2O$	2.0	2.7	3.3	16.6
$C_1 - C_3$ Gases	5.4	7.2	3.2	3.6
$C_3 - C_4$ Gases	8.8	11.8	3.0	3.1
C_5 - 200°C Naphtha	53.0	71.2	37.2	39.9
200 - 480°C Distillate	20.6	0	26.3	21.8
> 480°C	5.0	0	2.9	5.1
Total	94.8	92.9	75.9	90.1
Conversion, wt.% MAF Coal	92.8	90.2	76.7	85.8
Hydrogen Consumption, wt.% MAF	5.7	7.6	5.9	7.3

Source: Fuel, Vol. 60, Sept. 1981.

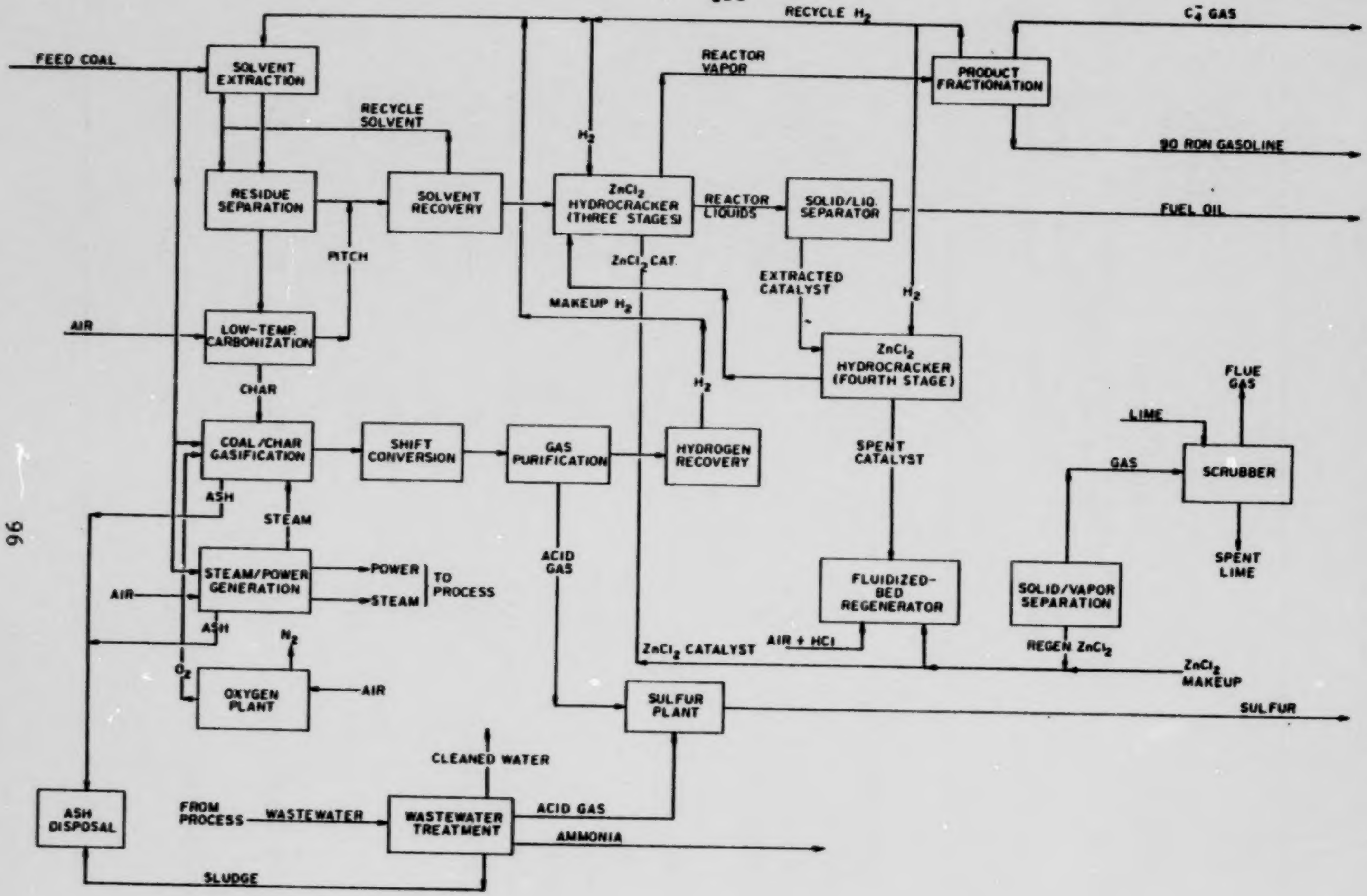

Figure 2-12 Block flow diagram for the Zinc-Halide Conceptual Commercial-plant.

Source: "Technical and Economical Evaluation of Selected Coal Liquefaction Processes", Oak Rigde National Laboratory ORNL/TM-7581, April 1981.

2.6.2 **Overall Conceptual Plant Configuration** (cont'd)

Hydroclone overflow is sent to a solvent recovery section and hydroclone underflow to a low-temperature carbonization step where additional liquid hydrocarbons are recovered. The hydroclone overflow consists of solvent and extract, but some residue remains entrained in this stream.

The low temperature carbonization (LTC) is carried out at 482 to 510°C and low pressure with the residue fluidized with air to form char. Some char is entrained overhead and is recovered as part of the pitch-solvent stream. Pitch and solvent from LTC step are sent to the solvent recovery block.

The bottom product from the solvent recovery step is the "extract" feed to the zinc chloride hydrocracking reactor. The hydrocracking reaction is carried out in a four-stage reaction system in order to achieve a high conversion to gasoline product. Between the third and fourth hydrocracking stages an extraction step provides for separation of unreactive oil from the recycle system. The gas oil distillate is used as extraction solvent which is recycled to the first stage with the extracted oil. The fourth hydrocracking stage also functions as a stripper to prevent the loss of product oil to the spent catalyst regenerator.

In the product fractionation block, the reactor vapors from three hydrocracker reactors are cooled, neutralized with caustic and sent to high pressure (27.2 MPa) phase separator. Off-gas from this separator is used as hydrogen recycle. The liquid is sent to a low pressure (1.7 MPa) phase separator where liquid

2.6.2 Overall Conceptual Plant Configuration (cont'd)

hydrocarbons separated from water phase are then sent to two distillation columns. The overhead material from the first column is directly fed to the second column which produces 90 RON, 10 RVP gasoline and LPG products.

The spent catalyst from the fourth stage hydrocracker reactor is sent to two fluidized-bed regenerators operating at 414 kPa and 982°C. A mixture of air and HCl gas are used to fluidize the beds. The overhead vapor from the regenerator passes through Mullite-lined cyclones. The cyclone overflow is condensed in a waste heat boiler and then fed to the gas-liquid separators. Liquid $ZnCl_2$ is withdrawn from the bottom of the separators and cooled to 316°C. Part of this liquid is recycled as regenerated catalyst and the remainder is used as regenerator quench liquid.

The char produced in the LTC unit is gasified to produce required hydrogen. The acid gases are sent to the sulfur plant for recovery of elemental sulfur as byproduct. The sour water streams from the entire plant are collected and stripped to recover ammonia.

The plants' fuel requirements are met by burning coal. Cooling water requirement are 325 L/s and electricity requirement of 55 MW are met by generating power within the plant.

2.6 ZINC-HALIDE PROCESS (cont'd)

2.6.3 **Technical Review**

2.6.3.1 Types of Coal Investigated

Wyoming Colstrip subbituminous coal (25% moisture and 12.5% ash) has been extensively tested in bench-scale units. Few runs were carried out with Rosebud Seam subbituminous coal from Montana. Amongst bituminous coals, only Kentucky No. 9 has been tested in batch autoclaves. In addition to these coals, extensive testing of coal extracts from Kentucky Nos. 9 and 14 produced in the SRC-I pilot plant has been tested. The PDU has been operated only with the SRC extract (0.3 wt% ash). A total of about 164 hours were logged with a longest uninterrupted run of 16 hours. Run lengths are defined as time during which organics and catalyst are fed to the reactor. The PDU has run 32 total hours on Neville solvent feed, 28 total hours on SRC-I diluted with Neville solvent, and 88 total hours on SRC-I diluted with recycled process-derived solvent.

One limitation of the process is that it cannot tolerate coals with high alkali content. As indicated earlier (see Section 2.6.1) the alkali salts lower the catalytic activity of zinc chloride. The total liquid yields reduce to 14 wt% MAF coal in the presence of 12.6 wt% Nacl in the feed compared to 35.3 wt% liquid yields when no salt was added in the feed. Table 2-17 presents the alkali content of different rank coals. The data indicate that Saskatchewan lignites have higher alkali content (3.95 wt% for Estevan and 4.66 wt% for Willow Bunch) compared to 3 wt% in the Wyodak Coal. The PDU runs have not been carried out with lignitic coals. Therefore quantitative assessment of reactor yields with lignites cannot be made at present.

Table 2-17 Alkali Salt Content in Different Rank Coals

Coal	Subbituminous	Lignites			
	Wyodak	Indian Head	Big Brown	Estevan	Willow Bunch
Ash Content (wt% MF Coal)	9.7	10.2	16.4	15.4	21.0
Alkali Salt Content (wt% MF Coal)					
CaO	2.33	1.63	2.16	2.32	3.0
MgO	0.55	0.50	0.60	0.57	1.05
K_2O	0.05	0.82	0.08	1.0	0.34
Na_2O	0.07	0.01	0.11	0.06	0.27
Total	3.0	2.96	2.95	3.95	4.66

2.6.3.2 Product Yields, Distribution and Efficiencies

The product yields and distribution of a conceptual commercial plant operating with Wyodak subbituminous coal are presented in Table 2-18. These results are based on 25,000 t/d 'as-received' coal and represent the net yields after taking the allowance for hydrogen requirements. About 47% of coal (8465 t/d) is utilized in the offsites to meet plant fuel, steam and electricity requirements. The results indicate that about 3.9 barrels of total liquid fuel (gasoline and fuel oil) is produced per tonne of dry coal to the reactor. However, based on 18,000 t/d dry coal fed to the plant, the liquid yield is 1.9 bbl/t MF coal.

The overall and liquid product efficiencies are 66.1% and 42.4% respectively. If the propane and butane LPG's are included, the liquid product efficiency increases by 4.3%.

2.6.3.3 Product Quality

The main liquid products from the Zinc-Halide process are C_5-200°C light distillate (gasoline) and fuel oil. The elemental analyses of these products are shown in Table 2-19. The sulfur content is low due to the inherent property of zinc chloride to act as a sulfur acceptor. The very low content of nitrogen is due to presence of zinc chloride which forms a complex with basic nitrogen compounds. The chloride content in the liquids is about 45 ppm.

The light distillate a produced in the process has a research octane number (RON) of 90.6 to 92.3. The motor octane number is 78.7 - 81.5. It passes the copper strip corrosion test and is

Table 2-18 Estimated Product Yield and Efficiencies in the Zinc Halide Conceptual Commercial Plant

[Basis: 25,000 t/d As-Received Coal]

	Wyodak
Coal (As-Received)	
Moisture, wt.%	28.0
Ash, wt.% of MF Coal	7.1
Coal Fed to Reactor	
Moisture, wt.%	-
Ash, wt.% of MF Coal	7.1
Input	
MF Coal to Reactor, t/d	9535
MF Coal to Offsites, t/d	8465
Electricity, MW	-
Output, t/d (bbl/d)	
SNG	1725
C_3LPG	232 (2621)
C_4LPG	199 (1949)
Gasoline (90 RONC)	2780 (23301)
Fuel Oil (200 - 480°C)	1825 (11153)
Sulfur	71
Phenol	-
Ammonia	68
Liquid Product Yield (C_5-480°C)	
bbl/t As-Received Coal to Reactor	2.6
bbl/t MF Coal to Reactor	3.6
bbl/t MAF Coal to Reactor	3.9
Gasoline in Total Liquid, wt%	60.4
Thermal Efficiency, %	
Overall	66.1
Liquid Products	42.4

Source: R. Salmon et al., "Technical and Economical Evaluation of Selected Coal Liquefaction Processes", Oak Ridge National Laboratory Report ORNL/TM-7581, April 1981.

Table 2-19 Distillate Analysis of Zinc-Halide Products

	Whole Distillate (C_5 - 480°C)	Naphtha* (C_5 - 200°C)	Mid-Distillate (200 - 325°C)	Heavy-Distillate (325 - 480°C)
Elemental Analyis (wt.%)				
Carbon	88.35	85.99	89.41	90.92
Hydrogen	10.82	13.47	10.19	8.71
Oxygen	0.8	0.52	0.30	0.22
Nitrogen (ppm)	(200)	(107)	(204)	(122)
Sulfur	0.01	0.01	0.08	0.14
PNA Analysis, Vol.%				
Paraffins		13		
Naphthenes		58		
Aromatics		29		

* melts gasoline specifications.

Source: M. Pell et al., 14th Intersociety Energy Conversion Conference, Boston, Aug. 5-10, 1979.

2.6.3.3 Product Quality (cont'd)

low in nitrogen (6.6 ppm) and sulfur (3.5 ppm). The PNA analysis is also within the gasoline range.

From the elemental analysis stand point, the distillate streams are comparable with fuel oil Nos. 2 and 6.

2.6.3.4 Status of Research and Development

The process development work was carried out under a cost-shared program by Conoco Coal Development Company, Shell Development Company and the U.S. Office of Coal Research. The work began in January 1975 which aimed at developing the process using zinc chloride. The work was carried out in laboratory, continuous bench-scale units, and in a 1-t/d PDU. Due to lack of funds from U.S. DOE and private organizations, the work on the project was terminated in late 1979.

Many aspects of the process have yet to be demonstrated. For example, the catalyst regeneration section of the PDU was only operated once. This step has only been demonstrated in batch autoclaves.

The reactors in a commercial plant are visualized as sparged tube columns, agitated by gas. These have not yet been demonstrated.

One of the main technical difficulties encountered are the corrosive products which can be formed due to the presence of zinc chloride. There is considerable uncertainty regarding the materials of construction, in particular the metals and ceramics

2.6.3.4 Status of Research and Development (cont'd)

used in the high-temperature reactor zone, in the zinc chloride condenser in the catalyst recovery system, and any other areas exposed to moist hydrogen chloride.

Unique environmental problems are expected from the disposal of solid materials containing high concentrations of zinc compounds.

snc

2.7 SOLVENT REFINED LIGNITE PROCESS

The Solvent Refined Lignite (SRL) process concept is being developed by the Department of Chemical Engineering at the University of North Dakota, U.S.A. The main objective is to investigate the liquefaction behaviour of lignites in the presence of syngas ($CO + H_2$) instead of pure hydrogen gas and to evaluate if this concept can be adapted in the Solvent Refined Coal (SRC) process for producing SRL product. The ash-free SRL product is solid at room temperature but can be separately hydrogenated to produce liquid hydrocarbons.

The liquefaction concept and approach is essentially the same as for the CO-Steam process described earlier in Section 2.2.

2.7.1 History and Development

The lignite liquefaction research at the University of North Dakota (UND) began in 1965 in so-called "cold charge" batch autoclaves. This work was initiated under the sponsorship of the Great Northern Railway Company. In 1972, the Office of Coal Research sponsored a $3.0-million contract for five years. During this period, the initial research was expanded through designing, construction and operation of a 23-kg/h process development unit (PDU) under the R & D program called "Project Lignite".

Pre-screening investigations were carried out in a one-litre capacity batch autoclave called "cold charge" autoclave. The experiments were carried out by charging the autoclave with slurry containing approximately 400 grams of hydrogenated anthracene oil (HAO 29), an equivalent of 200 grams of MAF coal, and an amount of water to give 0, 50, or 100 g of total water (corresponding to "dry", "half-dry", or "wet" coal). The

2.7.1 **History and Development** (cont'd)

autoclave was sealed, flushed with reactant gas, and then pressurised to 8.3 MPa (1200 psig) with hydrogen or syngas (equimolar H_2 and CO). After an overnight of constant stirring, the autoclave was heated electrically to a typical reaction temperature of 420° to 460°C. When the desired temperature was reached, the heaters were turned off and the reactor allowed to cool. This was considered to be the "zero time". When the autoclave had cooled to 205°C, the vapors were drawn off slowly through a series of cold traps. The dry gas leaving the traps was collected and analyzed. Materials in the traps were weighed and separated into oil and water fractions.

The autoclave was further cooled overnight, pressurized the next day with nitrogen, and the product slurry withdrawn through a dip tube and weighed. Residual slurry was also recovered from the autoclave by solvent washing. Elemental analysis for C, H, N, and S data based on solubility in tetrahydrofuran (THF) and cyclohexane (CH) were determined for the slurry. An ASTM D1160 vacuum distillation was performed at 667 Pa, which was carried to the smoke point or to the point at which the vapor temperature fell. The fractions of material in the IBP-120°C, 120-260°C, and 260°C-maximum temperature range (called light, middle, and heavy oil, respectively) were calculated. THF insoluble material less ash was designated insoluble organic matter (IOM). The yield of solvent-refined-lignite (SRL) was calculated as the difference between the distillation residue and the THF insolubles.

2.7.1 **History and Development** (cont'd)

The batch autoclave studies were aimed at the following items to determine the liquefaction behaviour of lignites:

- Effect of prolonged storage (70 weeks) of lignite in distilled water, air, and nitrogen before liquefaction. Results under this study revealed that samples stored in air gave about 87% conversion whereas the fresh lignite and those stored under nitrogen or water had higher reactivity which was about 92%. Air oxidation appears to cause lower reactivity.

- Effect of particle size of feed lignite between minus 100 mesh to 6.4 mm did not show significant change in the product yields suggesting that fine pulverization of lignite is not essential.

- Eliminating the moisture content from the lignite-solvent slurry before liquefaction adversely influenced the extent of liquefaction (24 - 42 wt% MAF unconverted lignite in the case of pre-dried slurry compared with 8 - 10 wt% unconverted lignite in the baseline tests).

- Liquefaction in the presence of syngas having H_2/CO mole ratio of 1 results in about 90 wt% MAF conversion compared with 80 wt% when either H_2 or CO is used alone. These results correspond to lignites tested under "as-received" conditions.

- The carbonization of lignite between 370 to 395°C before liquefaction results in lower (38 wt% versus 68 wt% MAF coal) net liquid yields.

2.7.1 **History and Development** (cont'd)

The University of North Dakota has also operated a 23 kg/h process development unit (PDU) capable of producing about 7 kg/h of solvent refined lignite (SRL) along with some hydrocarbon liquids and gases.

A simplified flow diagram of PDU is shown in Figure 2-13. In the slurry preparation area, pulverized lignite is continuously mixed with solvent in the ratio of 1:2. The resulting slurry is fed to the liquefaction reactor using high pressure reciprocating pump.

The slurry is mixed with synthesis gas containing 1:1 mole ratio of H_2 and CO before preheating and fed to the tubular upflow reactors. The heat in the preheater is provided by air-fluidized hot sand at about 427°C. Two reactors (5.5 m x 1.2 m) are used. The inlet of each reactor is at the bottom and the outlet at the top with another outlet provided at the center. This arrangement allows to change the reactor residence time by fourfold without changing the feedrate.

The liquefaction reactor effluent is separated by conventional high and low pressure flashes. The gases containing unreacted syngas, CO_2, H_2S and C_1 to C_3 hydrocarbons are sent to the gas recovery area where acid gases are first removed through absorption using sodium hydroxide and then through ammonia absorber containing boric acid solution. Part of the purified gas containing mostly syngas and some hydrocarbon gases is recycled along with makeup syngas.

The product slurry stream is sent to the solids precipitation tower in the solids-liquid separation area. The solids precipitation tower is essentially a settling tower in which feed

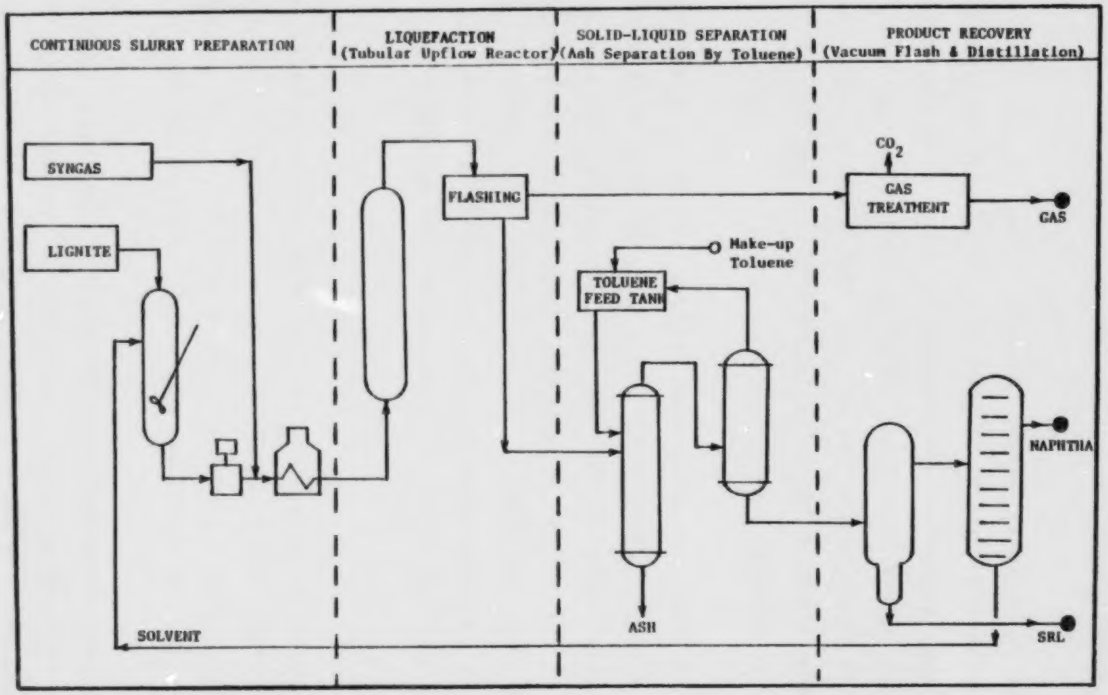
CONTINUOUS SLURRY PREPARATION
LIQUEFACTION (Tubular Upflow Reactor)
SOLID-LIQUID SEPARATION (Ash Separation By Toluene)
PRODUCT RECOVERY (Vacuum Flash & Distillation)
SYNGAS
LIGNITE
FLASHING
CO_2
GAS TREATMENT
GAS
Make-up Toluene
TOLUENE FEED TANK
NAPHTHA
ASH
SOLVENT
SRL

Figure 2-13: Simplified Diagram of 23 kg/h SRL Process Development Unit

2.7.1 **History and Development** (cont'd)

is dissolved in freshly added toluene. The ash and unreacted lignite thus settled at the bottom are withdrawn intermittently along with some toluene. The liquid overhead from the settling tower consisting of dissolved SRL, liquefaction solvent, and toluene is depressurized, cooled and then fed to the toluene recovery tower where toluene is flashed off. The bottom stream is sent to the product recovery area.

In the product recovery area, the product stream is first vacuum flashed at a pressure of 69 to 103 kPa. The bottom stream from the vacuum flash unit is the low ash and sulfur containing SRL product which is solid at ambient temperature. The overhead stream from the vacuum flash unit is fractionated and a stabilized to produce naphtha, the fractionator bottom is the liquefaction solvent which is recycled to the slurry preparation unit.

It must be pointed out that the PDU was never operated on a continuous basis. For example, the solids-liquid separation system using toluene could not be operated in an integrated manner due to several problems which are discussed later.

The PDU has been operated for four years. The main objectives were to identify and determine the effect of process variables on the liquefaction yields. These variables include reactor residence time, temperature, pressure, solvent-to-coal ratio, and CO/H_2 mole ratio in the feed gas. The range of these variables are:

2.7.1 **History and Development** (cont'd)

pressure	-	10.3, 13.8, and 17.2 MPa
temperature	-	360 to 480°C
slurry hourly space velocity	-	0.66 to 2.8 $m^3/h/m^3$
solvent-to-coal ratio	-	1.3 to 3.0
CO/H_2 mole ratio of feed gas	-	0.33, 1.0, and 3.0

The liquefaction response (expressed as conversion of coal on MAF basis) at 17.2 MPa and 370°C for North Dakota 'Indian Head' lignite is found to be 80 wt% which increases to 93 wt% if temperature is raised to 460°C. However, total conversion to naphtha and SRL product is maximum between 438 and 449°C. The conversion drops to about 76 wt% when liquefaction pressure is reduced to 10.3 MPa from 17.2 MPa. Thus, optimum conditions for both product (naphtha + SRL) yields and effective solvent balance are 444°C and 17.2 MPa.

The yields of light hydrocarbon gases and CO_2 increase with increasing temperature but decrease slightly with increasing pressure. The highest CO_2 yield was 68 wt% when the CO/H_2 mole ratio in the syngas was 3 to 1.

Increasing H_2/CO mole ratio in the feed gas increases the overall yields and more naphtha is produced. However, under this situation, though large amount of liquid product is available for recycle, the recycle solvent becomes lighter and consequently becomes less effective. For effective solvent recycle, CO/H_2 mole ratio of 1.0 is found as optimum.

2.7.1 **History and Development** (cont'd)

The Solvent Refined Lignite (SRL) product is a non-distillable but tetrahydrofuran (THF) soluble residue which is solid at room temperature. The average molecular weight of SRL is in the range 400 to 600 indicating that this material is composed of highly polycyclic type compounds. The catalytic hydrotreating of SRL permits its conversion to a liquid fraction which can be used as petroleum feedstock to produce transportation fuels.

The SRL hydrotreating studies were initially carried out in batch autoclaves using commercially available Ni/Mo, Co/Mo, and Ni/W catalysts supported on alumina. Later, some synthesized catalysts such as MoO_3-$NiSO_4$-TiO_2(x), MoO_3-$NiSO_4$-ZrO_2 of different sizes and physical properties (i.e., surface area, pore volume, acid content) were employed. The synthesized catalysts were prepared using nickel sulfate instead of usually used nickel nitrate. The purpose of nickel sulfate was to generate strong acid sites by the interaction between titanium oxide and sulfate anion which were expected to facilitate the hydrocracking of SRL.

The feedstock for hydrogenation was obtained from PDU runs using North Dakota lignite liquefied at 17.2 MPa and 400°C under an equimolal mixture of CO and H_2. The experiments were carried out for 2 hours reaction time at two different temperatures (400 and 450°C) and at three different pressures (6.9, 12.1, and 17.2 MPa). The hydrotreating results indicate that:

- at 400°C and 17.2 MPa, about 66 wt% liquid, 20 wt% gases and 14 wt% vacuum bottom residue yields can be obtained

2.7.1 **History and Development** (cont'd)

- increasing temperature to 450°C increases gaseous product yields by 50%

- overall conversion is poor (about 82 wt%) at 6.9 MPa. At pressures of 12.1 to 17.2 MPa, conversion increases to about 93 wt%

- the activities of synthesized catalysts were higher (30% with $Mo\text{-}NiSO_4\text{-}TiO_2$) when compared with commercially available catalyst

2.7.2 **Overall Conceptual Plant Configuration**

The overall conceptual process configuration of an integrated commercial plant has not been developed. For preliminary cost estimates, a conceptual configuration of four major plant sections (liquefaction, filtration, gasification and, oxygen production) was prepared in 1973 by the Ralph M. Parsons Company.

2.7.3 **Technical Review**

2.7.3.1 Types of Coal Investigated

Various lignites from geographically separate regions have been tested in batch autoclaves and a bench-scale unit. These lignites include two from Gulf Coast region, Texas ('Big-Brown' and 'Bryan') and five from the Northern Great Plains, North Dakota ('South Beulah', 'Indian Head', 'Larson', 'Velva' and 'Gascoyne'). The ultimate and proximate analysis of these lignites is presented in Appendix A. The principal difference in these lignites studied was in their ash composition, with the Northern lignites containing appreciable alkali content, especially sodium.

In addition, lignites from Denver Basin, Thailand, and Chile have also been tested. No significant difference in the liquefaction behaviour of all these lignites was observed with an exception for lignite from Chile. The Chilean lignite responded somewhat differently as judged from the 63% lower gaseous product yields when compared with other lignites (although the liquid yields were similar with other lignites). This difference appears to be related to the reduced utilization of inherent moisture for a source of hydrogen, probably because the minerals in the Chilean lignite lack the ability to catalyze the water-gas-shift reaction to the extent exhibited by the minerals in the other lignites.

'Indian Head' North Dakota lignite, which contains only 7 wt% ash 'as-received' basis, has been the most extensively tested coal.

Comparing the coal analysis data for Saskatchewan Estevan and Texas Big Brown lignites show that these two lignites are quite similar. With respect to the proximate analysis, Estevan lignite contains about 5 wt% less ash and about 7 wt% more moisture when compared with Big Brown lignite. These differences, however, will not cause any significant difference in terms of liquefaction response of Estevan lignite.

The Willow Bunch lignite from Saskatchewan is found to have similar analytical data corresponding to South Beulah lignite from North Dakota. On MF basis, the volatile matter is the same

2.7 SOLVENT REFINED LIGNITE PROCESS (cont'd)

2.7.3.1 Types of Coal Investigated (cont'd)

(40.6% versus 39.6% in Willow Bunch). The fixed carbon is about 4 wt% less and the ash content is 4 wt% more in Willow Bunch when compared with South Beulah lignite. Concerning the elemental analysis, Willow Bunch has 5 wt% higher oxygen and 1.7 wt% less sulfur. The significance of these differences suggest higher syngas or CO consumption for Willow Bunch lignite in liquefaction.

2.7.3.2 Product Yields

Typical product yields for 'Indian Head' North Dakota lignite are presented in Table 2-20. These are gross reactor yields expressed as wt% MF lignite. Also included in this table are the estimated liquid yields from hydrotreating of 'SRL' product. Thus the third column indicates the estimated total product yields in the SRL process which incorporates catalytic hydrotreating of SRL to produce maximum liquid fuels.

These results indicate the total naphtha and fuel oil yield of 42.3 wt% on the MF basis. This value of yield corresponds to about 50 wt% on MAF basis (29.5 wt% moisture, 10.7 wt% ash). The C_1 - C_3 hydrocarbon gas yields are about 17 wt% MF coal.

As a result of SRL hydrotreating, about 9.8 wt% ash-free, low-sulfur solid fuel (vacuum bottoms) product is also produced.

For a conceptual self-sufficient plant, all the solid fuel along with some hydrocarbon gases will be required to meet the plant fuel and make-up syngas requirements.

2.7 SOLVENT REFINED LIGNITE PROCESS (cont'd)

2.7.3.3 Product Characteristics

Product characterization studies have not been carried out.

2.7.3.4 Status of Research and Development

The R & D work on Solvent Refined Lignite (SRL) concept carried out by the Department of Chemical Engineering at the University of North Dakota could be considered as support research and development directed for conversion of lignitic coals using SRC advanced coal liquefaction process (which is mainly being developed for converting high-sulfur-content bituminous coals).

The 23-kg/h PDU at University of North Dakota was dismantled in 1977. The U.S. Department of Energy budget cuts have suspended the further development work.

Table 2-20 Typical Reactor Yields in the SRL Process

	Liquefaction	SRL Hydrotreating	
Process Conditions			
Temperature, °C	441	399	
Pressure, MPa	17.2	17.2	
Solvent-to-Coal Ratio	2.16	-	
H_2 in Feed Gas, mole %	74.2	100	
Overall Yields, wt% MF Coal			Total Yields*
CO_2	36.2	-	36.2
C_1 - C_3 Gases	8.2	9.1	17.3
Naphtha	12.2	-	12.2
Fuel Oil	-	30.1	30.1
SRL	49.0	3.4	9.8
Vacuum Bottoms	-	6.4	
Unreacted Coal	7.8	-	
Ash	8.6	-	
Sub-total	122.0	49.0	
CO Consumption	15.6	-	
H_2 Consumption	1.0	Not available	
H_2O Consumption	5.4	-	
Total	100.0	49.0	

* Sum of liquefaction and SRL hydrotreating yields from 'Indian Head' Lignite.

Source: D.E. Severson, A.M. Souby and J.J. Harris, "Process Development of Lignite Liquefaction", Proceedings of a Symposium on Technology and Use of Lignite", pp. 159-192, Grand Forks, North Dakota, May 18-19, 1977.

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CHAPTER - 3

COAL PYROLYSIS

3.0 COAL PYROLYSIS

Two pyrolysis processes were selected by the "Study Commitee" for assessment, particularly with respect to their readiness for commercialization and their overall suitability for the designated Saskatchewan lignites to produce liquid fuels.

The processes chosen were: (i) Lurgi-Ruhrgas and (ii) Occidental Pyrolysis. The technical review of the Lurgi-Ruhrgas process has been presented in Volume II along with other coal liquefaction processes considered to be at the advanced development stage. In this chapter, a review of recent developments in Occidental pyrolysis process is presented. This process is at an early stage of development and is not expected to reach commercial readiness before 1995.

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3.1 OCCIDENTAL PYROLYSIS PROCESS

The Occidental Pyrolysis process involves rapid pyrolysis of crushed coal particles at about 600°C in an entrained stream of hot coal, char, and a gas. The pyrolysis reactor vapors are quenched and stabilized by a hydrogen-rich donor solvent (called polystop-liquid) to produce a lighter liquid product than that obtained from indirect cooling of vapors. The main products are char and liquid hydrocarbons. A distinctive feature of the process is the use of the product char as a carrier to supply heat directly to the pyrolysis reactor in a continuous flow system. The other feature is the very short reactor residence time (few seconds) which allows higher liquid yield and higher coal throughput.

3.1.1 **History And Development**

Occidental Research Corporation (formerly Garret Research and Development Co.) a wholly owned subsidiary of Occidental Petroleum Corporation, U.S.A. initiated the development of this process in 1969 based on the concept of flash pyrolysis. Development started with a review of existing pyrolysis data which indicated that the most attractive route for an efficient conversion must incorporate high heat transfer (about 4000 kJ/m^2 hr °C) and short residence time (between 0.5 to 2.0 seconds). Mathematical model studies, with emphasis on short residence time parameters, demonstrated that pulverized coal can be heated to decomposition temperature in fractions of a second. These theoretical results were confirmed by experiments on highly volatile bituminous coals in a small, semi-continuous reactor having a nominal capacity of 2.3 kg/h feed coal. The operating experience with this reactor unit led to a number of improvements such as increased liquid yields, reduced reactor fouling problems, and to produce higher quality liquid products. In

3.1.1 **History And Development** (cont'd)

1976, a 2.7-tonne-per-day Process Demonstration Unit (PDU) was built and operated for two years. The two-year PDU program was sponsored by the U.S. Department of Energy (DOE).

During the earlier development of the Occidental Pyrolysis process, extensive use has been made of a small semi-continuous bench-scale reactor unit of 2.3-kilogram-per-day capacity. The key features of this unit are shown schematically in Figure 3-1. The system consists of a coal feeder, pyrolysis reactor, and char and tar collection equipment. Coal (-74 μm) is introduced by a Vibra screw feeder to the reactor inlet and pneumatically transported up the jet-mixer reactor in dilute-phase flow. Product char is removed by cyclones. Water-cooled condensers are used to collect tar. The heat to the reactor is provided by preheated char. The char preheating is carried out electrically in a separate fluid-bed char bin using nitrogen as fluidizing gas. The bench-scale results with highly volatile bituminous coal indicated a maximum conversion of coal to liquids of 35% at 1070°C (about twice the yield predicted by Fischer Assay and significantly higher than for fluidized-bed pyrolysis). The process variables identified to achieve maximum liquid yields were reactor residence time, pyrolysis temperature and char-to-coal feed ratio. Also the role of gases such as CO_2 and H_2O used to transport char to the reactor (instead of N_2) was investigated. The range of process variables investigated with both, bituminous and subbituminous coals are given below:

Reactor Residence Time	-	0.3 to 4.5 seconds
Pyrolysis Temperature	-	550 to 700°C
Char-to-coal feed ratio	-	10 to 40

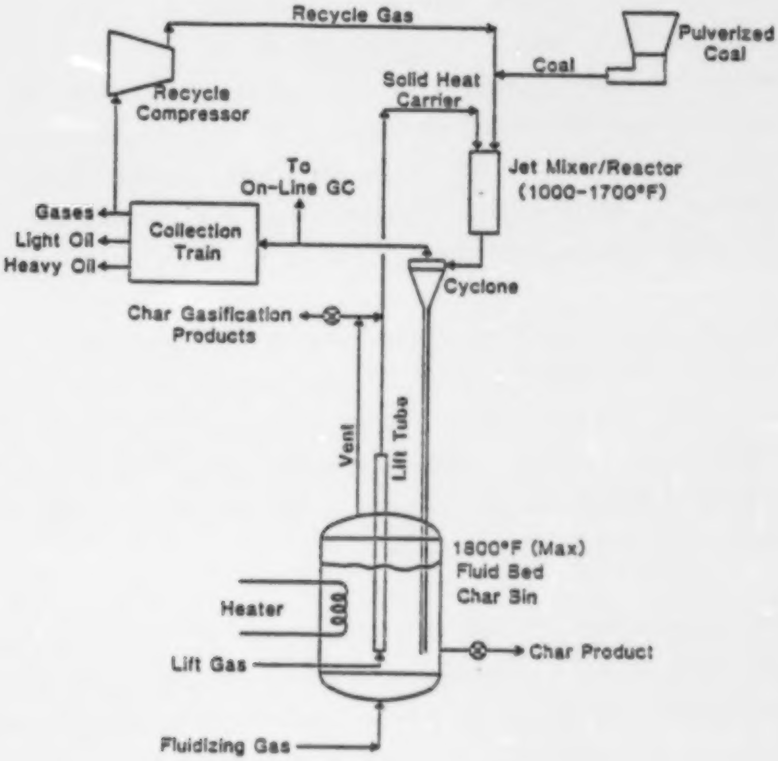

Figure 3-1 Schematic Diagram of 2.3 kg/h Occident Pyrolysis Bench-Scale Unit

Source: K. Durai-Swamy and W. Deslate, "Occidental Flash Pyrolysis Coal Liquefaction Process ands its Application to Power Generation", EPRI Conference on Coal Pyrolysis, Palo Alto, California, Feb. 25-26, 1981.

3.1 OCCIDENTAL PYROLYSIS PROCESS (cont'd)

3.1.1 **History And Development** (cont'd)

The maximum liquid yields, expressed as wt% on dry and moisture-free coal, is found to be 38 wt% at 607°C and 0.8 seconds residence time with bituminous coal. However, at 660°C and 2.0 seconds residence time, the liquid yield for the same coal was about 34 wt%. In the case of subbituminous coal, maximum yield of 22 wt% is attained at 1.4 seconds residence time and 649°C temperature.

At the same temperature, increasing residence time beyond 1.5 seconds, the liquid yields for both types of coal decrease. It is found that liquid yields also decrease with increasing char-to-coal weight ratios during pyrolysis under an inert gaseous atmosphere. By employing CO, CO_2, H_2O or CO_2+ H_2O gases, the liquid yields are enhanced.

The main objectives behind the two-year operation of the 2.7-t/d PDU were to obtain steady state heat and material balances, to demonstrate pyrolysis of caking coals without pretreatment and to evaluate hydrotreating of primary tar liquid to synthetic oil.

Three baseline runs were conducted to demonstrate the operability of PDU over extended period (over 120 hours) while obtaining reproducible material balance data. The first run yield data indicated 10 wt% tar (MAF basis) which was low compared to bench-scale data. Therefore, the objective of the second run was to maximize tar yields. The third run was aimed at demonstrating the maximum tar yields under maintained steady-state operation. The results of these runs are presented in Table 3-1. The tar yield results of all three baseline runs were nearly the same but uniformly lower when compared with the bench-scale results.

Table 3-1 Pyrolysis Yields in the Occidental Process

Process Unit	Bench-Scale		PDU		
(Yield Period Run)			121-77	122-77	123-77
Coal Type	Bituminous	Subbituminous	Subbituminous		
Coal Analysis, wt%					
Moisture	N.A.	-	18.5	13.9	15.0
Ash	N.A.	-	5.0	5.7	5.1
Volatile Matter	N.A.	-	33.8	36.1	35.8
Fischer Assay Tar (MAF)	17.0	12.0	11.6	10.6	10.0
Process Conditions					
Pyrolysis Temperature °C	580	580	594	649	649
Char-to-Coal Ratio	5.0	5.0	8	9	9
CO_2/CO mole Ratio	-	-	2.1	4.4	8.3
Product Yields, wt% MAF Coal					
Char	56.7	53.0	52.3	67.9	58.8
Tar	35.0	25.0	10.2	8.3	8.0
Gas	6.5	13.0	24.9	19.5	24.2
Water	1.8	9.0	12.6	4.3	9.0

Source: DOE Report FE-2244-14

3.1.1 **History And Development** (cont'd)

Factors which may have contributed to low yields included oxidation of the feed coal, leakage of combustion air into the pyrolysis zone, inadequate sampling procedures, and thermal cracking. A series of additional tests were structured for the PDU and bench-scale reactor to systematically evaluate these factors.

- Tar Hydrotreating

Tar hydroteating was carried out in a high-pressure autoclave unit attached to the PDU. Tar obtained from pyrolysis of Wyodak subbituminous and Western Kentucky bituminous coals were hydrotreated at about 17.2 MPa (2500 psig) pressure and 400°C temperature using pre-sulfided Co/Mo on alumina catalyst. The results of tar hydrotreating are discussed in pyrolysis product quality section.

3.1.2 **Overall Conceptual Plant Configuration**

A simplified flow diagram of the Occidental pyrolysis conceptual commercial plant is shown in Figure 3-2.

In the coal preparation block, the coal feed is dried during the crushing step by contact with hot-char combustion gases and then conveyed to the pyrolysis reactor by the recycle gas (containing H_2O, CO_2, CO, H_2, CH_4 and other light hydrocarbons).

In the pyrolysis block, the entrained-flow pyrolysis reactor operates at low pressure (up to about 520 kPa) with a ratio of recycle char to fresh coal of 5:1 or higher in order to heat the

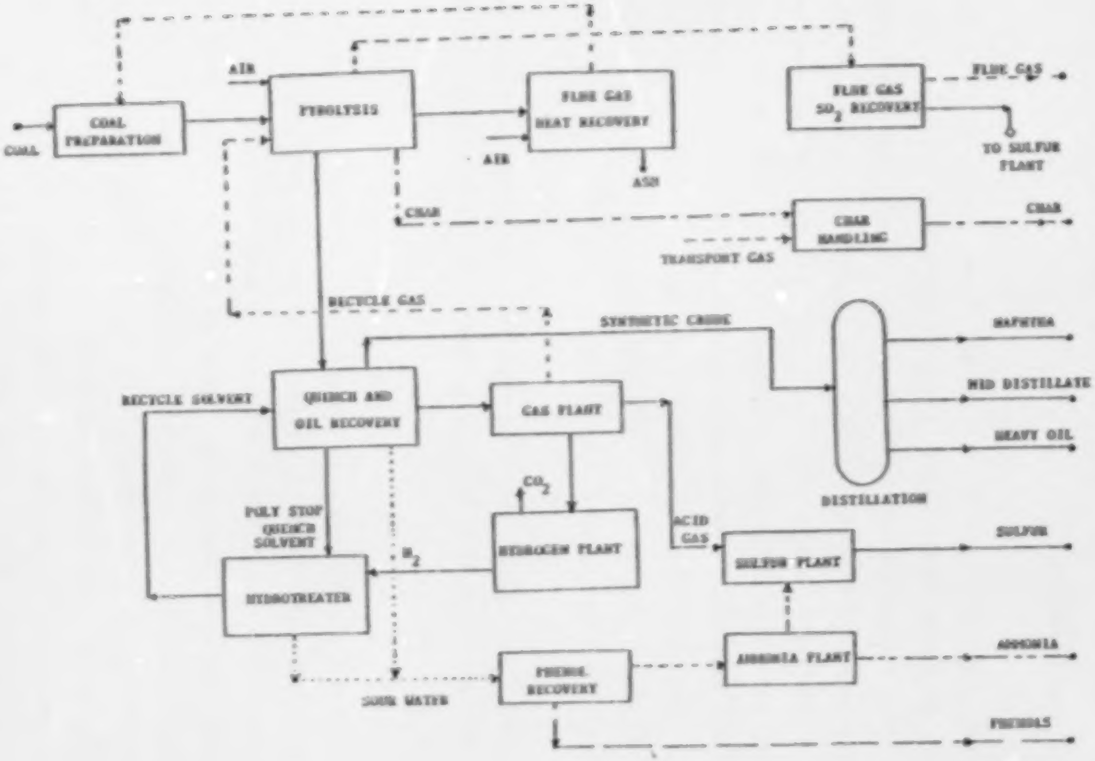

Figure 3-2 Block Flow Diagram of Occidental Pyrolysis Conceptual Commercial Plant

Source: K. Durai-Swamy and W. Deslate, "Occidental Flash Pyrolysis Coal Liquefaction Process and its Application to Power Generation", EPRI Conference on coal Pyrolysis, Palo Alto, California, Feb. 25-26, 1981.

3.1.2 **Overall Conceptual Plant Configuration** (cont'd)

feed to the reaction temperature of 580 - 600°C within a tenth of a second. The char and the reactive gases produce nascent hydrogen in the reactor.

The nascent hydrogen stabilizes the free radicals in the volatilized products minimizing their polymerization and thus, minimizing the formation of heavy tar and coke. The pyrolysis products are separated from the char by series of cyclones. The separated liquid and vapors are sent to quench and separation block. The hot char from the cyclones is collected in a surge bin. Part of this char is fed to a second entrained-flow vessel where it is partially combusted with air, raising its temperature to about 760°C. It is then recycled to the reactor after separation of the flue gases.

The flue gases are first sent to the heat recovery system and then recycled to transport coal from the coal preparation block to the pyrolysis reactor.

In the quench and separation system the product vapors are rapidly quenched by direct contact with a process-derived hydrogen-donor-solvent called 'polystop quench solvent'. This type of quench system apparently reduces the extensive product polymerization typical of pyrolysis products and results in a lighter liquid product stream. The quench separation system mainly consists of two liquid recovery stages. In the first recovery stage, raw pyrolysis gas and vapors are quenched to about 100°C condensing most of the vapors. Any entrained-char is also removed at this stage. The non-condensed pyrolysis gas from the first stage is sent to the second recovery stage where water and light liquids are removed.

3.1.2 **Overall Conceptual Plant Configuration** (cont'd)

The non-condensible gases from the quench separation system are sent to the gas plant where after the compression and acid gas removal, part of the treated gas is recycled as a carrier gas for the fresh coal feed. The remainder of the gas is sent to the hydrogen plant.

In the hydrotreater, heavy tar from quench and separation section is hydrotreated to restore the hydrogen-donor-ability of the polystop quench solvent.

The condensed water from the quench and separation system is sent to a phenol recovery unit where mixed phenols are produced as a byproduct. The sour water streams produced in the entire plant are collected and sent to the sour water stripper for the recovery of ammonia.

The acid gases from flue gas desulfurization, gas plant and hydrotreater block are sent to a Claus sulfur recovery plant where elemental sulfur is produced as a byproduct.

The remaining units are wastewater treatment, product storage and other conventional offsite facilities (not shown in Figure 3-2) comprising a self-sufficient plant.

3.1.3 **Technical Review**

3.1.3.1 Types of Coal Tested

Subbituminous Wyoming coal and bituminous Kentucky coal have been tested only.

3.1.3.2 Product Yields, Distribution and Efficiencies

The pyrolysis process can be operated under a variety of conditions designed to favour production of either gaseous or liquid products, or to achieve particular char characteristics. It is generally recognized that the most desirable approach would be the one which results in conditions designed to maximize liquid products. A considerable amount of work has been undertaken to establish optimum operating conditions to achieve this objective. The results indicate that a pyrolysing temperature of 600°C meets these requirements.

In bench-scale unit tests, Occidental has achieved oil yields of 25% wt% MAF from relatively low ash (6 wt%) subbituminous coals. This is equivalent to 200% of the Fischer Assay* yield. Oil yields of 30 wt% have been reported for bituminous coals.

In the recent tests carried out by Occidental with high ash (27 wt%) and moisture (18 wt%) subbituminous coal from Heatburg, Alberta, oil yields of only 15 wt% MAF were achieved.

The total marketable products from a conceptual commercial plant would include oil, char, product gas, sulphur, ammonia and phenols. Table 3-2 shows a typical product slate and yields from occidental process matched to a 2000-MW power plant using a typical Western Canadian subbituminous coal as feedstock. A 20 wt% oil yield is assumed (MAF coal basis).

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*An analytical method used to predict pyrolysis product yields from sample coal in the laboratory. Actual product yields from bench-scale or pilot plant pyrolysis runs when compared with the Fischer Assay reflects the degree of effectiveness of pyrolysis process carried out in a pilot-plant or in commercial practice.

Table 3-2 Estimate Product Yield in the Occidental Conceptual Pyrolysis Plant

[Basis: 25,000 t/d As-Received Coal]

	Wyoming Subbituminous
Coal (As-Received)	
Moisture, wt%	11.7
Ash, wt% of MF Coal	5.9
Coal Fed to Reactor	
Moisture, wt%	11.7
Ash, wt% of MF Coal	5.9
Input	
MF Coal To Reactor, t/d	22,075
MF Coal to Offsites, t/d	-
Electricity, MW	-
Output, t/d (bbl/d)	
Fuel Gas	556
Light Oil (31.7°API)	2522 (16,646)
Medium Oil (10°API)	1512 (8640)
Heavy Oil (4.4°API)	785 (4324)
Char	10,464
Sulfur	96
Phenols	42
Ammonia	19
Liquid Product Yield (C_5- 400°C)	
bbl/t As-Received Coal to Reactor	1.18
bbl/t MF Coal to Reactor	1.34
bbl/t MAF Coal to Reactor	1.43
Naphta in Total Liquid, wt%	52.3

Source: K. Durai-Swami and W. Deslate, "Occidental's Flash Pyrolysis Coal Liquefaction Process and Its Application to Power Generation", EPRI Conference on Coal Pyrolysis, Palo Alto, California, Feb. 25-26, 1981.

3.1.3.2 Product Yields,, Distribution and Efficiencies (cont'd)

Based on the data in Table 3-2, about 1.18 barrels of oil per tonne "as-received" coal (1.43 bbl/t MAF coal) can be produced in the Occidental pyrolysis process.

3.1.3.3 Suitability for Saskatchewan Lignites

Process suitability must be viewed on the basis of desired end-products. If char is considered as a main product, then the process is suitable for most types of coal. On the other hand, if liquid products are desired as the main products, a preliminary judgement on the process suitability may be made on the basis of Fischer Assay tar yields of the feed coal. For instance, coals with Fischer Assay tar yield less than 10%, as indicated by the process developer, should not be considered as suitable.

In Table 3-3, Fischer Assays of Estevan and Willow Bunch lignites are presented.* If a Fischer Assay tar yield of 10% is assumed for Estevan lignite, then the estimated amount of liquid products would approach 200 kg/t MAF coal (200% of the Fischer Assay). Char yield would be relatively much higher than liquid yields (see Table 3-2 for example). Therefore, the overall suitability of the process is related to the marketability of char. Char combustion in power plants may be viewed as one of the ways to utilize the char product.

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* The actual liquid yields from coal pyrolysis ranges between 120% to 200% of the Fischer Assay.

Table 3-3 Fischer Assay of Saskatchewan Lignites as Determined by B.C. Research Laboratories

[Carbonization Temperature = 500°C]

	Estevan Lignite		Willow Bunch Lignite
	Untreated	Predried to 10 wt% H_2O	Untreated
Fischer Assay Yield, wt% MAF coal			
Char	65.0	69.4	65.8
Tar	9.4	9.4	5.8
Water Formed	10.1	9.6	15.5
Total Gas (by difference)	15.5	11.6	12.9
Composition of Gas, vol%			
CH_4	17.5		13.8
C_2H_6	3.6		2.2
H_2	28.0		44.9
CO	9.0		4.2
CO_2	41.9		34.9
H_2S	0.1		0.1

Source: Liquefaction and Pyrolysis Tests of Samples of Estevan and Coronach Lignites prepared for The SNC Group, B.C. Research Laboratories Project Report 2-02-691, 1982.

3.1.3.4 Product Quality, Utilization and Upgrading Schemes

Properties of raw liquid (tar) produced from Wyoming subbituminous coal in the bench-scale and PDU runs are presented in Table 3-4. It can be seen that the nitrogen and oxygen content is exceptionallv high. The asphaltenes and preasphaltenes content is over 40% which lowers the quality of liquid far away from the conventional crude oil. In addition, high viscosity and tendency of tar repolymerization under prolonged storage prohibits its use as liquid fuel.

As indicated earlier, tar hydrotreatment has been carried out by the process developer. Typical results are presented in Table 3-5. The hydrotreating was carried out at 1.7 MPa pressure and 400°C temperature. Product inspection reveals a 27°API increase in specific gravity and over 95% removal of sulfur, asphaltenes and preasphaltenes and 75% removal of nitrogen and oxygen. This is achieved at the hydrogen consumption rate of 88 m^3/bbl. High conversion of preasphaltenes and asphaltenes to oils and sulfur removal can be obtained under moderate severity. Nitrogen removal will require higher severity.

The pyrolysis process also produces significant amounts of char and fuel gas as products. It is found that the heating value of the char slightly exceeds that of the parent coal despite a doubling of the ash content. This largely results from a corresponding reduction in moisture and oxygen content during the drying and devolatilization stages.

An important characteristic of the char is its sulfur content, approximately 63% of that in the parent coal, which would make this a compliance fuel under the recently issued National

Table 3-4 Properties of Pyrolysis Tar* from Wyodak Coal

Process Unit	Bench-Scale	PDU
Yield, wt% MAF coal	12.7	10.1
Gravity, °API	-8.1	-8.1
Viscosity, kg/ms @ 93°C	0.45	0.21
Distillation (ASTM-D1160)		
IBP °C	204	160
50% off	427	386
Non-Distillable, %	37	27
Composition, wt% (char-free)		
Carbon	79.06	79.09
Hydrogen	7.76	7.24
Nitrogen	1.04	1.48
Sulfur	0.39	0.45
Oxygen (by difference)	11.73	11.46
Characteristics, wt%		
Oil	56.4	59.6
Asphaltenes	27.4	28.6
Preasphaltenes	16.2	11.8

* Full range tar comprises the fraction with an initial boiling point of 110°F.

Source: DOE Report FE-2244-14, June 1977

3.1.3.4 Product Quality, Utilization and Upgrading Schemes (cont'd)

Guidelines for New Stationary Sources, Thermal Power Generation Emissions.

Chars are utilized in power boilers. Since the physical and chemical nature of chars vary considerably for different processes, it is difficult to generalize. Some chars are amenable to direct firing in conventional pulverized fuel firing systems whereas others would require further pulverization or use of fluidized-bed technology. Also, some chars are more reactive than others which require supplementary fuel addition to sustain ignition. Char arising from the Occidental pyrolysis process appears to be suitable for direct firing in conventional dry bottom pulverized fuel fired boilers.

Chars derived from Saskatchewan lignites which will have high inherent ash content would require particular attention in the design of boiler heat transfer surfaces to avoid prohibitive rates of fouling and/or abrasive wear. An increased load will also be placed on particulate removal and ash handling systems.

In the typical analysis of product gas, ethane and ethylene together account for about 15% of the constituents whereas higher hydrocarbons including aromatics account for a further 5%. The high content of CO (30%) will require the institution of rigid safety systems for personal protection similar to those employed within the petrochemical industry.

Table 3-5 Analysis of Hydrotreated Tar from Wyodak Coal

Feed : PDU Tar from Subbituminous Coal, Run 120
Catalyst : Co/Mo on Alumina, 1/16-in Extrudates
Temperature : 400°C
Pressure : 1.7 MPa

Experimental Results	Before Hydrogenation	After Hydrogenation
Specific gravity, °API	-7.6	17.0
Elemental Analysis, wt%		
Carbon	85.8	88.32
Hydrogen	6.9	10.40
Nitrogen	1.2	0.26
Sulfur	0.5	0.01
Oxygen (by diff.)	5.6	0.8
Chemical Analysis, wt%		
Asphaltenes	19.3	0.1
Preasphaltenes	10.3	0.57
Heteroatom Removal, wt%		
Nitrogen	-	79.3
Sulfur	-	96.0
Hydrogen Consumption, m^3/bbl	-	89

Source: DOE Report FE-2244-14, June 1977

3.1.3.5 Areas Requiring Further Developments

The Occidental pyrolysis process has not been operated on a large pilot-plant scale. Therefore, an assessment of specific problem areas at this stage cannot be carried out. During PDU operations, higher coal feed rates (91 kg/h) caused fouling due to erratic flow of hot char to the reactor. Mechanical failures particularly with valves were also the cause of PDU shutdowns.

3.1.3.6 Status and Commercialization Prospects

The Occidental Research Corporation (ORC) has been working on the pyrolysis process at the La Verne, California plant for 12 years. A 2.7-t/d PDU which operated between 1976 and 1978 has now been dismantled.

Two main research programs are in progress at ORC. One of them is a U.S. DOE sponsored three-year cost-sharing research program, which began in the last quarter of 1980. The estimated cost of this program is $4,56 million. One of the major objectives of the program is to understand the chemistry of flash pyrolysis reaction using Electron and Nuclear Magnetic Resonance Spectroscopies (EPR, NMR and ENDOR). This program is being carried out using a 2.3-kg/h bench-scale unit.

The second study was sponsored by EPRI. This involves assessment of the technical and economic features of Occidental's Flash Pyrolysis process and to assess its potential for producing clean fuels for the electric power industry. ORC has engaged Foster Wheeler Energy Corporation as a subcontractor to design the conceptual commercial plant processing 18,150 t/d coal and

3.1.3.6 Status and Commercialization Prospects (cont'd)

estimate its capital and operating costs. Because of cuts in DOE funds, the ORC coal pyrolysis research program will likely be terminated in 1982.

In Canada, the Occidental Pyrolysis concept is being examined for a project known as Alberta Electric Power Project (AEPP). This project is sponsored by Alberta Power Limited, Fording Coal Limited and Trans-Alta Utilities Corporation. Studies to determine technical and economic feasibility, for a 2,000-MW power plant integrated with Occidental Flash Pyrolysis is in progress. It is assumed that if the results of these feasibility studies provide sufficient encouragement then a demonstration unit of 850-t/d 'as-received' coal capacity will be required.

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CHAPTER - 4

REVIEW OF COAL/OIL CO-PROCESSING APPROACH

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4.0 REVIEW OF COAL/OIL CO-PROCESSING APPROACH

Direct coal liquefaction (hydrogenation) process in which process-generated recycle oil is replaced with an independently available heavy petroleum fraction or virgin petroleum oil, heavy oils or oil sands bitumen as a solvent may be referred to as the co-processing option of a direct coal liquefaction process.

The co-processing approach has a number of potential advantages over both the direct coal liquefaction (hydrogenation) and heavy oil/bitumen upgrading (hydrocracking). Some of these advantages are:

- reduction in molecular hydrogen consumption because oil-sands bitumen and heavy oil act as low-cost hydrogen-donor solvents
- reduction in capital cost because of less processing required, e.g., elimination of recycle system for coal-derived solvent for slurry preparation, reduction in make-up hydrogen production, etc.
- increase in liquid product yield over the processing of the bitumen or heavy-oil alone
- reduction in coke formation/deposition on the reactor walls.

The use of heavy oil for coal/oil co-procesing is being investigated in several R&D facilities around the world.

In Japan, the basic solvolysis process is being developed by Mitsubishi Heavy Industries (MHI), using petroleum-derived vacuum residue as a solvent. The process solvent includes a petroleum asphalt in a make-up component amounting to less than 25 wt% of the feed coal. The preheated coal/oil slurry is fed into a stirred dissolver/reactor operating at 400°C and atmospheric pressure. No molecular hydrogen is added in the dissolver. This is particularly attractive for processing of low rank coals because oxygen is removed without the consumption of hydrogen.

In the U.S.A., the Hydrocarbon Research Inc. (HRI) is developing the COIL process. Here, a virgin petroleum oil is used as a solvent. The coal/oil slurry is hydrogenated at high temperature and pressure in an ebullated-bed catalytic reactor.

In the USSR, there have been reports on research in co-processing of coal-aromatic heavy petroleum oil fractions.

In the Federal Republic of Germany, Veba Oel has carried out preliminary investigation of coal/heavy-oil co-processing concurrently with the development of VEBA-COMBI-CRACKING process aimed primarily for upgrading of heavy oil. The exploratory tests made by Veba Oel revealed that coal/heavy-oil co-processing option is feasible and could result in reduction of overall energy consumption of an integrated plant as compared to using coal-derived solvent (direct coal liquefaction). However, low content of paraffins and high aromatics and naphthenes in the heavy-oil is critical for coal/oil co-processing.

Canadian research using bitumen as a solvent was initiated as early as 1935. Recently, at the Alberta Research Council, Edmonton, an autoclave test program using Cold Lake bitumen, Lloydminster heavy oil and other heavy oil fractions derived therefrom is being carried out to evaluate co-processing option.

4.0 REVIEW OF COAL-OIL CO-PROCESSING APPROACH (cont'd)

Other Canadian laboratories (Raylo Chemicals, Sandwell Beak Research Group, University of Regina) under contract to CANMET are also investigating co-processing in an autoclave. The results have shown that the liquid yield for co-processing is about 34% higher than the thermal hydrocracking of bitumen alone. At higher hydrogen partial pressures (7 to 7.5 MPa), about 85% of the "maximum" liquid yield can be obtained. Maximum liquid yield refers to that obtainable if all the coal were converted to benzene-soluble liquid. With optimization of operating conditions it may be possible to increase yield to 95%.

Process development work on co-processing has also been carried out at CANMET since 1975. It has been shown that co-processing of coal/heavy-oil/bitumen could prevent the formation of coke on the reactor walls during hydrocracking of bitumen. This phenomenon is explained by the fact that the coke formed during thermal hydrocracking of bitumen is accumulated on the coal particles rather than on the reactor walls. Also, a substantial amount of coal was dissolved under bitumen hydrocracking conditions.

Thus, co-processing of coal and heavy-oil has a number of potential merits, especially for the unique situation in the province of Saskatchewan where both the coal and heavy-oil resources are available. However, up till now, the co-processing concept has been investigated mostly at the autoclave level and many problem areas have yet to be resolved on bench-scale and PDU units before demonstration on a pilot-plant level and further commercialization of the process.

In the following section, the two co-processing alternatives - basic MHI Solvolysis and HRI COIL processes which have been tested on bench-scale units is presented.

4.1 COIL PROCESS

In the COIL process, one part of coal or lignite is mixed with about two parts of virgin petroleum oil. The resulting slurry is hydrogenated in an ebullated-bed catalytic reactor at elevated temperature and pressure (400-510°C, 10.3 - 17.2 MPa). The reactor effluents contain gaseous hydrocarbons, heavy synthetic oil and unreacted coal and ash. The heavy oil is separated from phenolic compounds and is then further hydrogenated in a second-stage ebullated-bed reactor. The final products are light naphtha and distillate fuel oil. The residue slurry containing unreacted coal and ash is gasified to produce make-up hydrogen for the process.

The extent of conversion in the first-stage reactor can be varied. Lower conversion would yield more char which can also be used to produce fuel gas for the plant. Alternatively, severe operating conditions in the first-stage reactor may be employed in which case the second-stage hydrogenation step may be eliminated. The liquid product is highly aromatic and has a boiling range of heavy petroleum crude oil. This primary product may be fractionated to produce naphtha and heavy fuel oil. If lighter hydrocarbons are desired, then second-stage hydrogenation is necessary.

4.1.1 History and Development

The COIL process was developed and patented (U.S. Pat. 3,755,137) by Hydrocarbon Research Inc. (HRI) in 1973. The COIL process is based on the application of the commercially proven ebullated-bed reactor (originally developed by HRI to convert heavy oil residue into lighter fractions) for coprocessing of coal and oil.

4.1.1 **History and Development** (cont'd)

The process concept has been tested only in a bench-scale unit. Illinois No. 6 bituminous coal was used in these tests. Four cases were experimentally examined. Case 1 consisted of feeding coal without oil. In case 2, equal parts of coal and oil were fed to the hydrogenation reactor, whereas in case 3, twice as much oil as coal was used. Case 4 consisted of feeding the crude oil only to the reactor. The details of processing conditions such as pressure, temperature, amount and type of catalyst used as well as the size of the reactor used are not available. The analysis of feed coal and oil employed in the experiments is presented in Table 4-1 and the summary of the experimental results under each case is given in Table 4-2. A discussion of these results is presented later.

Table 4-1 Analysis of Feed Coal and Oil in COIL Process

Feed	Coal	Oil
Type	Illinois No. 6	Residuum Fraction (8° API)
Elemental Analysis, wt%		
Carbon	67.25	83.62
Hydrogen	4.81	10.29
Nitrogen	1.02	0.33
Sulfur	4.85	5.29
Oxygen	12.14	-
Ash	9.93	-
	100.0	98.3
ASTM Distillation		
IBP., °C		452
7.3 vol.%		524
Distilled, wt.%		6.3
Residue, wt.%		93.3
Loss, wt.%		0.4

Source: G.R. DeVaux et al., "COIL Process: First Generation Commercial Coal Liquefaction Plant", Miami International Conference on Alternative Energy Sources Proceedings, vol. 7, pp. 709-711, 1977.

Table 4-2 Summary of Experimental Yields and Product Quality in COIL Process

	Case 1	Case 2	Case 3	Case 4
Feed composition, wt%				
Coal	100	48	32	0
Oil	0	52	68	100
Experiemental Yields, wt% Dry Feed				
CO_x	0.8	0.23	0.10	0
C_1-C_3 Gases	6.8	4.78	4.04	3.5
C_4-200°C Naphtha	15.0	12.65	13.44	15.4
200 - 525°C Distillate	36.7	46.56	52.77	50.1
524°C+ Residue	18.7	21.00	18.12	28.3
Unconverted coal	3.8	2.87	1.68	-
Total	81.8	88.08	80.15	97.3
Hydrogen (Consumption)	3.6	3.0	2.8	-1.7
Product Analysis				
C_4-200°C Naphtha				
Gravity, °API	57.00	63.10	62.20	60.00
Sulfur, wt%	0.07	0.07	0.03	0.15
200-345°C Distillate				
Gravity, °API	15.00	25.70	29.60	34.00
Sulfur, wt%	0.07	0.23	0.09	0.63
345-525°C Distillate				
Gravity, °API	-2.00	11.70	15.60	20.00
Sulfur, wt%	0.20	0.83	0.50	1.35
525°C Residue				
Gravity, °API	-20.00	-3.60	-4.00	5.00
Sulfur	0.6	2.43	2.00	4.00
Benzene Insoluble, wt%	15.5	0.18	0.08	0.10
200°C Liquids				
Sulfur, wt%	0.27	1.14	0.74	2.12

Source: G.R. DeVaux et al., "COIL Process: First Generation Commercial Coal Liquefaction Plant", Miami International Conference on Alternative Energy Sources Proceedings, vol. 7, pp 709-711, 1977.

4.1.2 Overall Conceptual Plant Configuration

Figure 4-1 presents an overall conceptual plant configuration of the COIL process. In the preparation block, coal is dried to remove all surface moisture. The drying operation is carried out in the range of 140 to 371°C temperature in order to remove easily destroyable oxygen in the form of CO, CO_2 or H_2O. The dried coal is ground to average 100-mesh size, and then is slurried with crude petroleum or heavy oil. About 5 to 100 grams of cobalt-molybdenum catalyst per tonne of dry coal is added in the coal-oil slurry. The resulting slurry is preheated in the range of 400 to 510°C and mixed with hydrogen gas at about 13.8 MPa before it is fed to the first-stage liquefaction reactor.

The reactor is operated in the range of 400 to 510°C temperature and at about 10.3 to 13.8 MPa pressure. The reactor space velocity is in the range of 240 to 2403 $kg/h/m^3$. The relative size of the coal and catalyst particles and condition of ebullation are such that the catalyst is retained in the reactor while the unreacted char is carried out with the reaction products.

The reactor effluents are separated into a gas and a slurry stream. The slurry stream is passed through a coal-oil separation unit from which a heavy synthetic crude is removed from the char (unreacted coal). Heavy synthetic crude containing phenols and cresols is sent to a phenol recovery unit. The non-phenolic crude is sent to a second stage hydrogenation reactor. This reactor is also an upflow ebullated-bed reactor similar to the one used in H-Coal or H-Oil process.

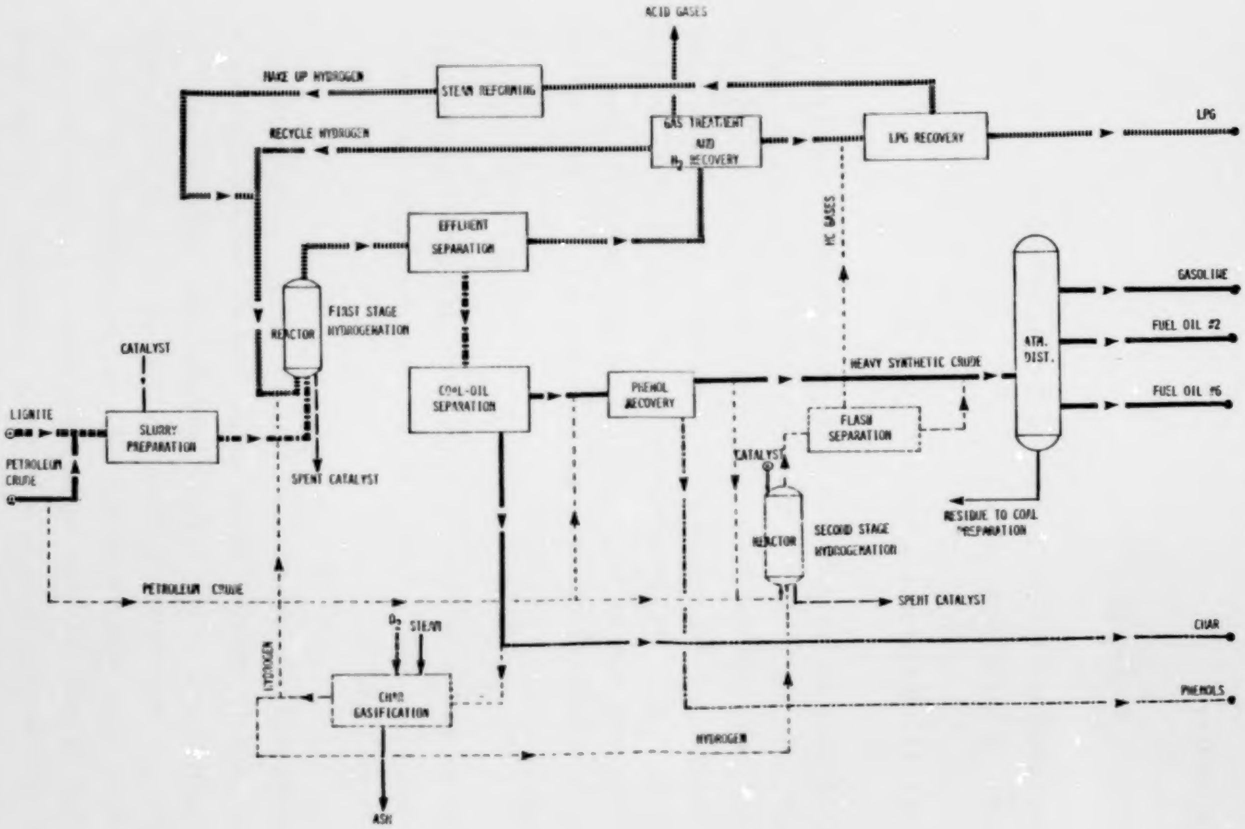
4.1.2 **Overall Conceptual Process Configuration** (cont'd)

The effluents from the second-stage hydrogenation reactor are separated into a hydrocarbon gas stream and a raw liquid product stream using flashing techniques.

The raw liquid product is fractionated (near atmospheric pressure) to yield gasoline (95 CFRR octane rating), a No. 2 distillate of about 32° API gravity, and a No. 6 fuel oil of 0° API gravity. The residual fraction from the fractionator is utilized as part of the recycle oil to prepare coil-oil slurry. The non-condensed gases from the first-stage reactor effluent separation are treated to remove H_2S, CO_2 and COS in the gas plant block. The treated gas is then combined with the hydrocarbon gases from second-stage hydrogenation and fed to the LPG recovery unit where mixed butanes as LPG product and a light hydrocarbon gas stream primarily containing C_1/C_2 is produced.

The C_1/C_2 gas is steam reformed to produce make-up hydrogen required for the first-stage hydrogenation. The char recovered after first-stage hydrogenation is gasified to produce make-up hydrogen for the second-stage hydrogenation.

All the acid gas streams containing H_2S and CO_2 produced by the entire plant are sent to the Claus sulfur plant (not shown in the Figure 4-1) for recovery of sulfur as byproduct. Ammonia is also recovered from sour-water streams through the ammonia recovery unit.



Notes

----OPTIONAL PROCESSING
Second Stage Hydrogenation Combined with Petroleum Crude Improves the Quality of Gasoline Product

Figure 4-1 Overall Conceptual COIL Process Configuration

Source: U.S. Patent 3,755,137

4.1.3 **Technical Review**

4.1.3.1 Types of Coals Investigated

As stated earlier, the COIL process has been tested only with Illinois No. 6 bituminous coal.

4.1.3.2 Product Yields, Distribution and Efficiencies

The experimental yields for four cases investigated were presented earlier (see Table 4-2). The C_4-200°C fraction has a higher API gravity and less sulfur in COIL than either the all-coal or all-oil cases. Between the results of case 2 and 3, higher yield of C_4-525°C liquids is attained when the feed consists of one part of coal and two parts of oil. Also the yield of residual oil (> 525°C material) is lower in this case when compared to all other cases.

Inspection of sulfur content in the 200°C+ products reveals that less sulfur (0.74 wt% versus 1.14 wt%) is present in case 3 in comparison with case 2.

As indicated earlier, the COIL process is simpler than the H-Coal process. This is because there is no need to recycle the coal-derived slurry oil, the entire system for recycle oil is eliminated. In the H-Coal process, approximately 2.75 kg of feed is required for each kilogram of liquid product produced. Because the recycle oil must be depressurized, repressurized, and reheated, the overall thermal efficiency is reduced. In the COIL process, both the oil and the coal are passed through the reactor once. The yield is about one kilogram of product per

4.1.3.2 Product Yields, Distribution and Efficiencies (cont'd)

kilogram of feed. Thus the thermal efficiency in the COIL process would be higher compared to the efficiency achieved in the H-Coal process. For bituminous Illinois No. 6 coal, the "overall" thermal efficiency of a conceptual self-sufficient, commercial plant H-Coal is 70.7% and the "liquid" product efficiency is 61.9%. For COIL process the overall and liquid product efficiencies based on the data presented in Table 4-3 are estimated as 72.8% and 71.4% respectively.

4.1.3.3 Process Suitability

The process has not been investigated beyond laboratory level. Besides, the lower rank coals have also not been tested. Therefore, the process suitability with Saskatchewan lignites cannot be assessed. However, in a commercial plant setting, the COIL process will employ an H-Oil or H-Coal reactor. In view of this fact, the suitability of Saskatchewan lignites in COIL process may be assumed to be similar to the H-Coal process (See Volume I for details of H-Coal Process). High moisture and ash content is found as the main areas of concern in H-Coal process (for details see Volume I, Section 3.1.4).

The process inputs for a conceptual, self-sufficient H-Coal plant operating with Wyodak coal are 430 L/s raw water, 5.4 TJ/h fuel and 193 MW electricity based on 25,000 t/d of coal "as-received" to the plant. The amount of fuel gas produced within the plant is sufficient to meet fuel requirements and to generate required power. The only utility required from an outside source is about 430 L/s of raw water.

4.1.3.4 Product Quality, Utilization and Upgrading

The process developer has not made any specific investigation to evaluate product quality. Based on the process concept, technical details under this category will probably be similar to those in the H-Coal process. It may be noted that if the second-stage hydrogenation is included in the COIL process configuration, the hydrotreating severity in the upgrading step will be low.

4.1.3.5 Status of Research and Development

As indicated earlier, only a few bench-scale runs have been made using Illinois coal. Similar runs will be required using Saskatchewan lignites in order to judge the merit of this process relative to straight H-Coal or other direct liquefaction process. The process developer has also indicated similar views about the suitablity of their process for Saskatchewan lignits (Appendix B).

Recognizing the magnitude of efforts required for the further development of the COIL process, it appears that this process will not be available for commercialization before 1995. At present there are no funding commitments from the process developer or from an other organization such as U.S. Department of Energy concerning R & D on the COIL process.

Table 4-3 Estimated Product Yield and Efficiencies in the Conceptual COIL Plant

[Basis: 25,000 t/d As-Received Coal]

Coal	Illinois No. 6
Coal (As-Received)	
Moisture, wt%	10.4
Ash, wt% of MF Coal	9.93
Coal Fed to Reactor	
Moisture, wt%	N.A.
Ash, wt% of MF Coal	9.93
Input	
MF Coal to Reactor t/d	17,845
MF Coal to Offsites, t/d	4,405
Oil to Reactor, t/d	37,921
Output t/d	
Naphtha (C_5-200°C)	7,495
Fuel Oil (> 200°C)	29,428
Sulphur	2,872
Phenols	-
Ammonia	178
Thermal Efficiency, %	
Overall	72.7
Liquid Products	71.4

4.2 JAPANESE SOLVOLYSIS COAL LIQUEFACTION PROCESS

Japanese solvolysis coal liquefaction (JSCL) process converts coal into light and heavy distillate oils as main products. Methane-rich gas is produced as co-product. The main feature of the process is that it uses a vacuum residue (asphalt) obtained from crude petroleum to solubilize coal into asphaltenes or pre-asphaltenes. The other features of the JSCL process are:

- Pulverized coal is solubilized in absence of hydrogen in a non-catalytic reactor at about 400°C temperature. Asphalt and recycled pitch are used as a solvent.

- The reactor (dissolver) is operated at atmospheric pressure because of the absence of volatile solvent and hydrogen gas. The solubilized coal is separated into ashless solvolysis pitch and residue containing ash and undissolved coal.

- The pitch may be catalytically hydrogenated and then separated by distillation into light and heavy oil fractions and a heavy bottom residue.

- If the heavy products of pitch hydrotreating are used as the solvent, the integrated process may become independent of asphalt make-up.

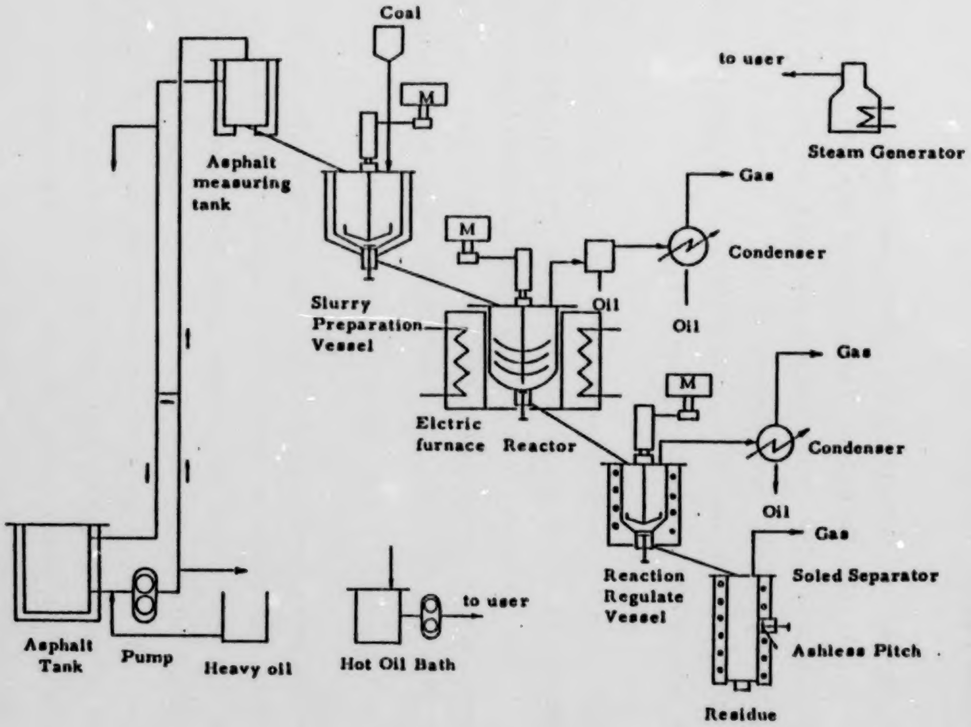
4.2.1 **History And Development**

In 1974, the Japanese government initiated a project called "Project Sunshine" to explore alternative sources of energy with greater emphasis on coal liquefaction and gasification. About 70% of the total $14.6-billion budget over the period from 1974 to 1990 has been allocated for coal liquefaction and gasification research and development.

4.2.1 **History And Development** (cont'd)

Initially, in the solvolysis coal liquefaction process only asphalt was used as solvent and the product was pitch like, a solid material at room temperature and containing all organic sulfur and nitrogen of the coal. Later, the process was modified and called "Solvolysis Phase II Process" in which the pitch is hydrogenated and then fractionated by vacuum distillation. The solvent consisted of vacuum distillation residue plus asphalt. The demonstration of the process has been carried out in laboratory scale units and in the 1-tonne-per-day process development unit (PDU). The pitch hydrogenation studies were carried out separately based on the TOKOSHE MDS process (developed in the study of hydrodesulfurization of heavy fuel oils) in batch autoclaves.

In the development of the modified solvolysis liquefaction concept, fundamental research was required in regard to the change in solvent properties due to recirculation of distillation residue with fresh asphalt. For this purpose, a laboratory-scale semi-continuous unit was built. A simplified flow diagram of this unit is shown in Figure 4-2. The unit consisted of batch reaction autoclaves connected in series. The unit was operated 8 to 10 times a day for solvolysis reaction test. The ash content of the ashless pitch was found to be between 0.5% and 0.6%. Other details such as percentage ash in the coal feed, oil-to-asphalt to-coal feed rate, properties of resulting oil-asphalt solvent are not available.



Figrue 4-2 Semi-Continuous Bench Scale Japanese Solvolysis Coal Liquefaction Unit

4.2.1 **History And Development** (cont'd)

A 1-tonne-per-day solvolysis plant (PDU) was commissioned in 1978 at the Nagasaki Technical Institute of Mitsubishi Heavy Industries Limited. The construction and part of the operating costs of this plant were funded by the Janapese Ministry of International Trade and Industry (MITI). A schematic flow diagram of the PDU is shown in Figure 4-3.

The Nagasaki plant commenced operations with Mi-ike coal and vacuum residue obtained from the refining of Iranian crude petroleum. The reactor feed was preheated elecrically to 370°C temperature to avoid sedimentation in the lines (in the commercial process, however, the feed slurry will be heated to 410°C).

An engineering development unit (EDU) was later built in 1979 with the main objective to establish a scale-up criteria for the design of a 40-t/d pilot plant. The EDU consisted of a reactor, solid discharge equipment and heater as shown in Figure 4-4. The coal supply chute (V-1) was designed nearly vertical to prevent clogging that had been observed during the operation of 1-t/d unit. The heater (F-1) employed was of once-through induction type in which the heating tube generated heat. A similar heater will be used in the 40-t/d plant. The reactor was a horizontal vessel having a double-shaft agitator. The design and configuration of this reactor was based on the operating experience with similar type reactor used in the 1 t/d unit. Various methods for discharging

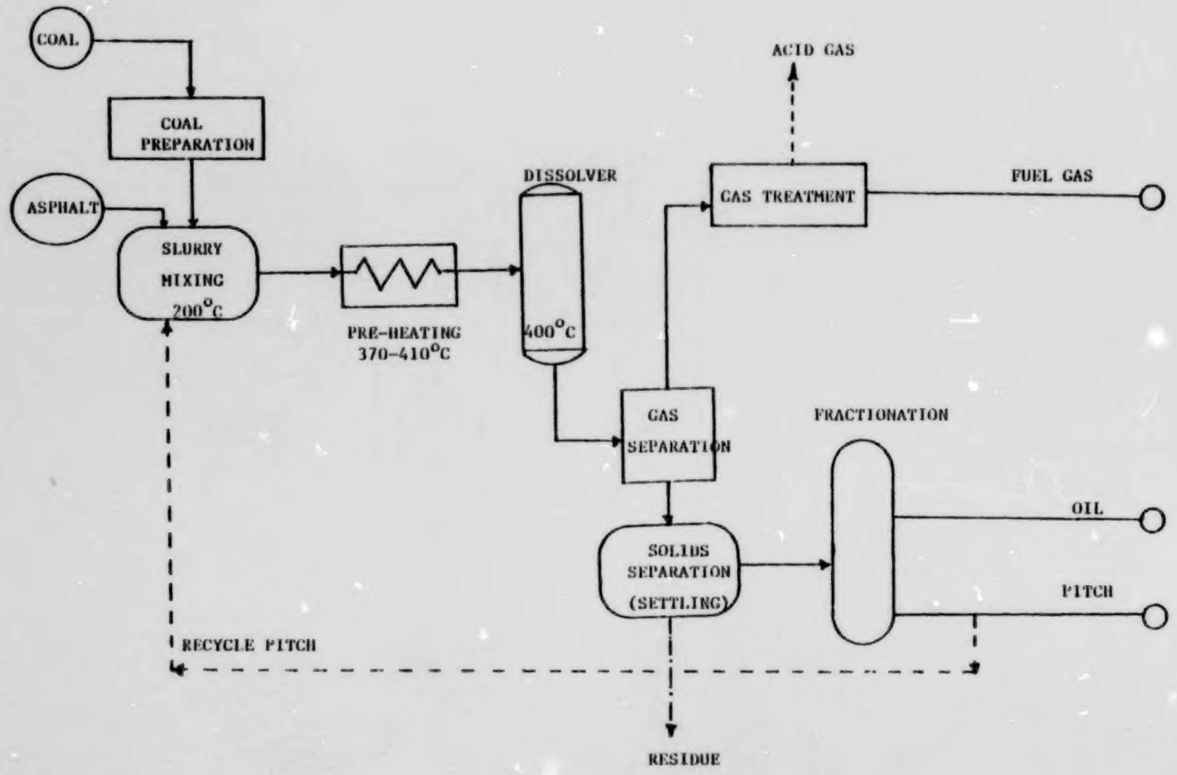


Figure 4-3 Schematic Flow Diagram of 1-tonne-per-day Nagasaki Pilot Plant

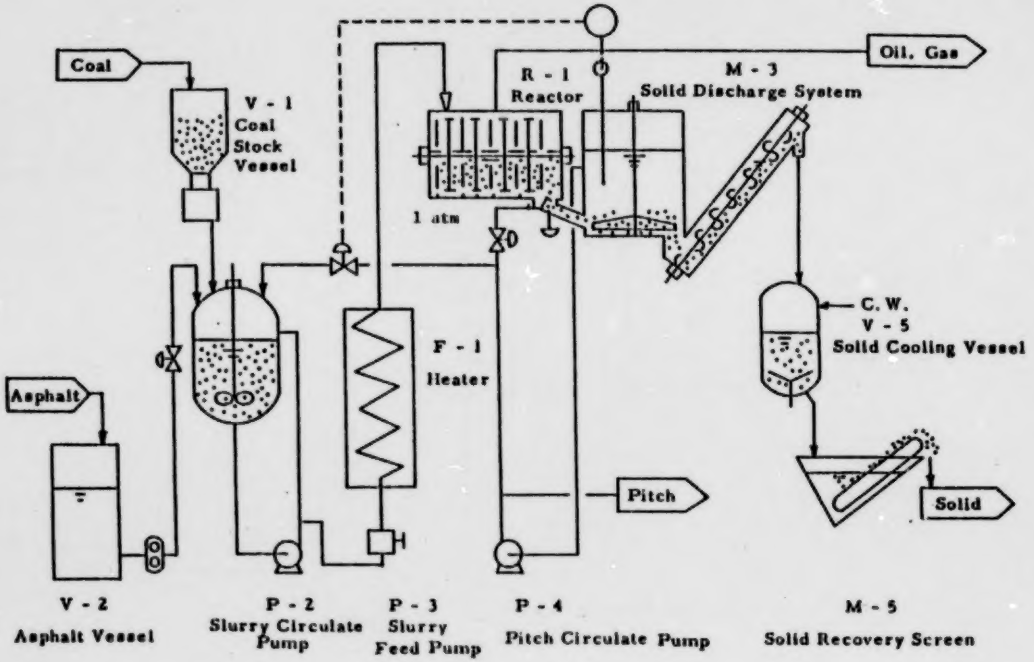


Figure 4-4 Schematic Flow Diagram of Engineering Development Unit for 40-t/d Japanese Solvolysis Coal Liquefaction Plant

4.2.1 **History And Development** (cont'd)

highly dense sedimented solid were evaluated including discharge by pumping, through a control valve or by a screw conveyor. The preferred solid discharge system (M-3) finally consisted of a vertical cylinder having a solids scraping mechanism for the sedimented solids and a screw-type solids discharging device.

- Solvolysis Pitch Hydrogenation Studies

These studies were carried out in 0.2,, 0.3, and 5 litre capacity batch autoclaves using Co/Mo and Ni/Mo as catalysts under the following operating conditions:

Reaction Temperature, °C	400
Total Pressure, MPa	19.6
Reaction Time, min	180
Catalyst/Pitch Ratio	0.1

Under above conditions, product oil obtained exhibited crude oil distillation properties rather than the properties of heavy oil. When temperature, catalyst-to-pitch ratio, and reaction time were increased, hydrogenation rates also increased and the product obtained was light oil. Increasing pressure from 14.7 to 19.6 MPa, however, did not improve the hydrogenation results significantly. The hydrogenation product properties were found different with pitches obtained from different coals. However, quantitative data, particularly the hydrogen consumption, oil yields and analysis are not available.

4.2.2 Solvolysis of Low-Rank Coals

The coal solubilization ability of asphalt is restricted to coals containing carbon from 80 wt% to 85 wt%. In fact, coals with carbon content lower than 80 wt% or higher than 85 wt% are difficult to solubilize when asphalt is used as a solvent. Lignitic coals, in general, have carbon content lower than 80 wt%. Therefore, in order to solubilize low rank coals, reactions of coal with olefins using asphalt as a solvent were carried out in the laboratory using batch autoclaves. The olefins investigated were ethylene and propylene. The experimental procedure and results are summarized below.

One part of pulverized Taiheiyo coal (76.3 wt% carbon) and two parts of Khafji asphalt were heated for 60 minutes in an autoclave in the presence of ethylene. The initial pressure of ethylene in the autoclave was 5.9 MPa. At the end of 60 minutes, the reacted mixture was extracted first with benzene. The benzene insolubles were then extracted with quinoline. The extent of solubilization was defined in terms of percentage solubility in benzene or quinoline. At 390°C, about 70 wt% of Taiheiyo coal was converted to quinoline soluble matter. When a similar experiment was carried out using propylene, about 55 wt% of this coal was converted to quinoline soluble matter. When no olefins were used, the quinoline soluble matter was only 35 wt%. These results indicate that solubilization of low-carbon coals must be carried out in the presence of ethylene. Further studies on the role of ethylene using model compounds confirmed that improved solubilization is due to two reasons. The first is alkylation of the aromatic nucleus by ethylene. The second is due to higher reactivity of ethylene with the organic oxygen present in coal. Based on these findings, it appears that Saskatchewan lignites, which have low carbon content (52 wt% to 62 wt% MF basis) and high oxygen content (17 to 22 wt% MF basis),

4.2.2 **Solvolysis of Low-Rank Coals** (cont'd)

would require reaction with ethylene to increase the solvolysis yields with asphalt.

4.2.3 **Overall Conceptual Plant Configuration**

A simplified conceptual plant configuration for the Japanese solvolysis liquefaction process is shown in Figure 4-5.

Pulverized coal (-015 mm) is mixed at about 200°C with vacuum residue solvent and petroleum asphalt in the ratio of 1:2. The amount of make-up asphalt added is about 25 wt% based on coal feed rate. The coal-solvent slurry is preheated to between 400 and 410°C and pumped to a stirred dissolver/reactor. Because of the absence of volatile solvent and hydrogen gas, the reactor is operated at around 207 kPa with residence time of 30 minutes. The solvolysis coal liquid from the reactor is sent to a separation unit where it is first quenched with asphalt and then passed through a vapor-liquid separator/condenser.

The condensed phase enters a settler where solids removal is assumed to be assisted by the formation of 'coke' in the reactor. The residue-containing ash is removed from the settler and sent to a hydrogen plant. The ashless solvolysis liquid (pitch) is hydrogenated and then is vacuum distilled to recover light and heavy oil products. The bottoms from the vacuum distillation unit are recycled back to the slurry preparation block.

The non-condensed gases from the dissolver and separator are sent to a gas treatment plant where methane-rich fuel gas product is separated from acid gases. The acid is sent to the sulfur plant (not shown in the figure) for the recovery of elemental sulfur.

4.2.4 Technical Review

Even though the Japanese solvolysis coal liquefaction process is advancing to a pilot-plant stage, information is not available in the open literature. Furthermore, on a non-confidential basis, material provided by the process developer is insufficient to assess the overall suitability of lignitic coals. Therefore, the technical review presented here is carried out on a preliminary basis.

4.2.4.1 Types of Coal Investigated

Japanese, American, Canadian and Australian coals ranging in carbon content from 66.3 wt% MAF coal (Morwell brown coal) to 90.1 wt% MAF coal (Smoky River) have been tested in laboratory-scale batch autoclaves. Yallowin and Liddel seam coals have been tested in a bench-scale solvolysis unit and other Australian coals including Blackwater, Lemington and Blair Athol have been studied in small batch units. Japanese Mi-ike coal has been processed in a 1-tonne-per-day process development unit.

4.2.4.2 Product Yields and Distribution

The relationship between product yield and carbon content of coals from 66.3 wt% (Morwell brown) to 90.1 wt% (Smoky River) are presented in Figure 4-6. The gas, oil (condensable distillate) and quinoline insoluble residue yields, as shown in this figure, were obtained after 60 minutes of solvolysis reaction with petroleum asphalt at 390°C temperature and near ambient pressure. Under these conditions, coals containing less than 80 wt% carbon (MAF basis) give a constant yield of about 20 wt% quinoline insoluble (QI) residue on MAF basis, thus indicating 80 wt% solvolysis or solubilization of coal feed. Coals of

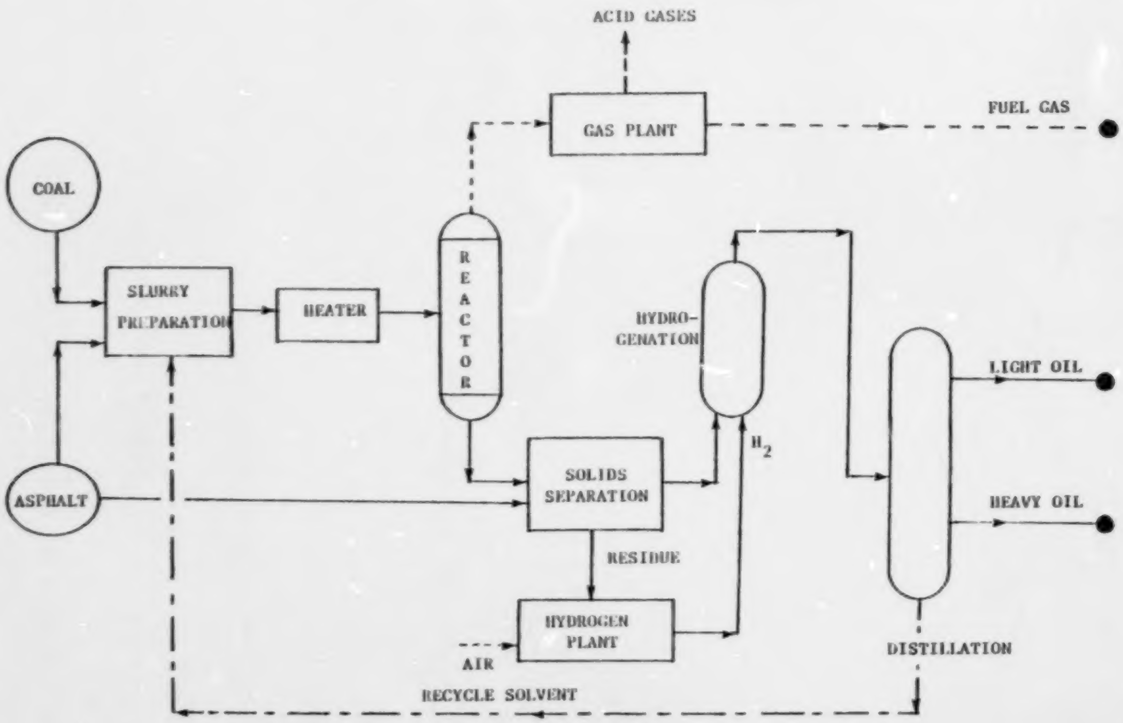


Figure 4-5 Overall Conceptual Japanese Solvolysis Process Configuration

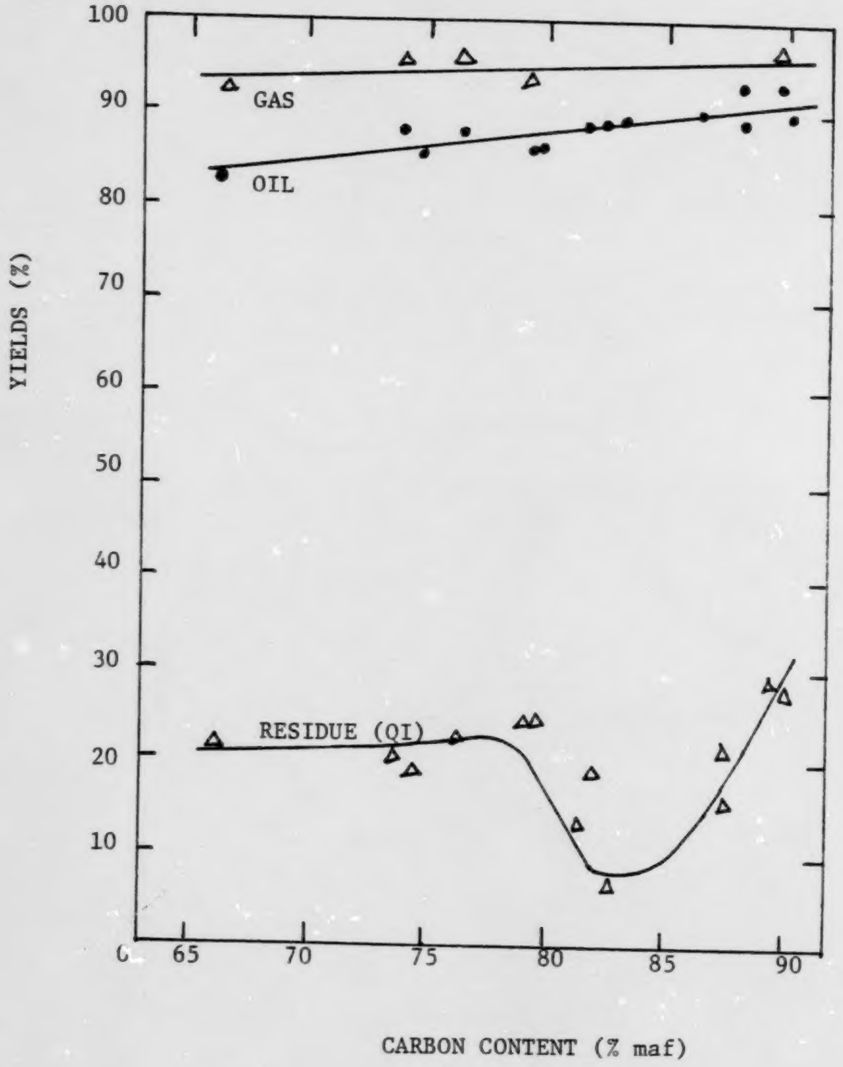


Figure 4-6 Solvolysis Product Yields

4.2.4.2 Product Yields and Distribution (cont'd)

80-86 wt% carbon give the lowest yields of QI residue. The minimum QI yield of 10 wt% is obtained from coal with 83 wt% carbon. Above 86 wt% carbon content, the yield of QI residue increases rapidly. Thus, low volatile prime coking coals such as Australian Beatrice and Smoky River are unsuitable for solvolysis.

The low conversion yields experienced with high rank coals appear to apply to a wide range of hydrogenation conditions. However, the conversion behaviour of the low rank lignites and brown coals appears to depend much more on the solvolysis conditions.

The Saskatchewan lignites contain 66 wt% carbon on MAF basis as in Estevan and 72 wt% carbon as in Willow Bunch. Based on the results shown in Figure 4-6 these lignites may give about 20 wt% (MAF) QI residue yields.

The overall material balance from the operation of 1-tonne-per-day Nagasaki pilot plant (with no pitch recycle to the solvolysis reactor) indicates about 38 wt% oil yields based on the following results:

Feed		Products	
Coal	100	Gases	10
Asphalt	235	Distillate	38
	335	Pitch	232
		Residual Solids	55
			335

4.2.4.2 Product Yields and Distribution (cont'd)

Hydrotreating of the pitch product to increase liquid yields may offer a solution to the reduction of surplus solid residues. A heavy recycle fraction from this hydrotreater may prove to have hydrogen donor properties which will increase conversion in the solvolysis reactor.

4.2.4.3 Process Suitability

Based on the concept of the Japanese solvolysis liquefaction process, it appears that this process may prove to be attractive for lignitic coals (which have high oxygen content) because oxygen is removed without the consumption of hydrogen. However, to achieve higher solvolysis yields, solvolysis of lignitic coals will have to be carried out in the presence of ethylene. Therefore, availability of ethylene has also to be considered in the case of processing lignites. Bench-scale tests must be carried out to determine the amount of ethylene required per tonne of lignite as these data have not yet been established.

4.2.4.4 Product Quality, Utilization and Upgrading

Details on this subject are not available in the open literature. It may be pointed out however, that the concentration of phenols (which are known to cause polymerization) in the solvolysis oils will be higher when compared with oils produced from direct coal liquefaction processes using hydrogen gas in the reactor.

4.2.4.5 Areas Requiring Further Research and Development Work

One of the major operating difficulties experienced so far has been to control the quenching of solvolysis reactor effluent. It is important to achieve an even distribution of the quench liquid in the effluent, otherwise the overquenched zones solidify causing massive blockages. This may prove to be difficult in a large scale operation.

The absence of molecular hydrogen during solvolysis (apart from the small amount of donatable hydrogen from asphalt) apparently leads to polymerization of primary products to produce the precursors of coke. In fact, coke formation in the reactor has been observed.

4.2.4.6 Status and Commercialization Prospects

The Nagasaki pilot plant, commissioned in 1978, was designed to process 1 t/d of coal. A larger unit, processing 7 t/d of coal, has been planned and this plant will include pitch recycle and possibly hydrotreating, which are not incorporated in the Nagasaki plant.

A major objective of the project is to acquire sufficient experience of coal conversion processing to allow for the evaluation of this technology.

Commercial viability of the Solvolysis Process, however, would depend on finding a suitable application of byproduct residue which will have approximately twice the ash of the feed coal.

4.2.4.6 Status and Commercialization Prospects (cont'd)

A great deal of development work has yet to be carried out for the design, construction and operation of a 40-t/d pilot plant. Studies on product characterization, utilization and upgrading with respect to end-use compatibility have to be carried out also.

The support for further developmnet of Solvolysis process has not been finalize yet. However, the Japanese Government has committed funds for the "Sunshine Project" till 1990, therefore, funding for the Solvolysis process are likely to be available.

CHAPTER - 5

SUMMARY - CONCLUSIONS AND RECOMMENDATIONS

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5.0 SUMMARY - CONCLUSIONS AND RECOMMENDATIONS

In previous chapters, the state-of-the-art review of direct coal liquefaction and pyrolysis processes and concepts and coal/oil co-processing options selected by the Study Committee was presented. Also, preliminary evaluation of indirect coal liquefaction processes in combination with three second generation gasifiers, Westinghouse, U-Gas and Shell was carried out by Foster-Wheeler as a part of the Study. The Foster-Wheeler report to SNC on this part of the Study in included in Annex A.

For a given coal rank and type, among the more important factors affecting the level of hydrogenation and the related yield and product distribution are the catalyst and the operating variables - pressure, temperature, and residence time. Listed in Table 5-1 are representative operating conditions, hydrogen consumption and yields for direct liquefaction processes, DOW, CO-Steam, CSF, TSL, SRC, Z-H and SRL. A summary of operating conditions and yields for Occidental pyrolysis and coal/oil processes is given in Table 5-2.

- Operating Conditions

  The operating temperature for DOW, CO-Steam, CSF, TSL, SRC, Z-H and SRL processes ranges from 415 to 460°C. The operating pressure for DOW and CO-Steam processes, which are one-stage processes, is the same (14 MPa). In the case of CSF, TSL, SRC, Z-H and SRL processes, which are two-stage processes, the operating pressures are different in the first-stage depending upon the method of hydrogen addition. In the CSF process,

Table 5-1 Reactor Operating Conditions, Hydrogen Consumption and Yields

Process[a]	DOW	CO-STEAM	CSF	TSL	SRC	Z-H	SRL
Reactor Conditions 1st Stage/2nd Stage[b]							
Temp., °C	460	460	480/430	460/415	440/415	415/415	440/400
Press., MPa	14	14	3/21	17/19	11/19	14/28	17/17
Res. Time, Min.[c]	N A	60	40	3	30	60	60
Yields, Wt.% MAF Coal 1st Stage/2nd Stage[b]							
C_1-C_4 Gases	19.6	15.6	13.7[f]	3/3	6	24.0[f]	9/10
Naphtha	15.7	51[d]	33[e,f]	3/8	9/15	31.4[f]	14
Mid-Distillate	23.5			3/39	7/17	20.6[f]	34[g]
Heavy Oil	12.5	20.4		16	2/4		
SRC				6			
H_2 Consumption, Wt.% MAF Coal	5.2	6.0[f]	6.0[f]	0.6/4.5	2.7/3.1	8.3[f]	5.7[h]/NA

Notes:

a) Feedstock - Bituminous coal for DOW, CSF, TSL and SRC. Subbituminous for coal Z-H and lignite for CO-Steam and SRL.
b) Where applicable.
c) For first stage (liquefaction) reactor only. Data not available for second stage reactor.
d) Represents naphtha plus distillate oil
e) Gasoline.
f) Combined data for first and second stage
g) Mid-distillate plus heavy oil obtained as a result of SRL hydrotreating
h) Includes hydrogen produced in-situ by water-gas shift reaction. Hydrogen equivalent CO of 4.2 wt% in CO-Steam and 4.6 wt% in SRL process.

Table 5-2 Summary of Operating Conditions and Yields in Pyrolysis and Coal-Oil Processes

Process	Pyrolysis	Coal-Oil	
	Occidental*	JSCL*	COIL*
Operating Conditions			
Temperature, °C	600	440	430
Pressure, MPa	0.1	0.2	13
Yields, wt% MAF Coal			
C_1-C_4 Gases	3[a]		
Naphtha	17	10	5
Mid-Distillate	10	-	15
Heavy Oil	-	62[b]	59[c]
Char	48		

Notes:

a) Net yields after tar hydrotreating

b) Coal extract - would require hydrotreating

c) Includes mid-distillate and heavy oil

* Feedstock - Subbituminous coal in occidental, Japanese Mi-Ike coal and JSCL and Illinois No. 6 bituminous coal in COIL process.

coal is first solubilized (extracted) in the presence of a process derived hydrogen donor solvent. Thus, hydrogen is added indirectly through the solvent, therefore, operating pressure is low (3 MPa). In the case of TSL, SRC, Z-H and SRL processes, molecular hydrogen is added directly from the gas phase. Therefore, operating pressure are high ranging from 11 to 17 MPa.

The pressure in the second stage, which is principally a vapor-phase catalytic hydrogenation for all two-stage processes, is in the range of 17 to 28 MPa. The different pressure employed in different processes is due to intrinsic selectivity related to product slate chosen. For example, in the Zinc-Halide process, the second stage is operated at 28 MPa pressure and the product slate consists of gasoline (90 RON clear) with a small amount of mid-distillate oil. The heavy oil fraction is not produced at all. In the case of the TSL process, the second stage is carried out in an ebullated-bed reactor at 19 MPa pressure and the product yield includes a maximum amount of mid-distillate oil and a relatively much lesser amount of naphtha and heavy distillate.

In the pyrolysis process, choice of temperature and pressure affects product amount and composition of volatile material. Occidental's experience with bituminous coals indicates 600°C temperature and ambient pressure can provide maximum amount of pyrolysis liquids (tar plus oil).

The feature of coal-oil processes called "co-processing" lies in the use of petroleum-derived solvent instead of coal-derived liquid. However, operating pressure and temperature in these processes is similar to those employed in direct coal liquefaction processes. For example, Japanese solvolysis process which is based on extraction concept (similar to CSF process), employs 440°C temperature and low pressure around 2.0 MPa (480°C and 3 MPa in CSF process). In the case of COIL process, since hydrogen is transferred from gaseous phase to the coal-oil slurry (made of 1 part coal and 2 parts petroleum oil), the operating pressure and temperature is 13 MPa and 430°C respectively, which lie in the range employed in direct liquefaction processes.

- Product Yields and Hydrogen Consumption

The results of product distribution and yields in Table 5-1 and 5-2 show that varying amounts of hydrocarbon gases (C_1-C_4), naphtha (C_5-200°C), mid-distillate (200 - 350°C), and heavy oil (350 - 520°C) are produced by various processes. It should be noted that boiling points indicated for each distillation fraction are typical but vary slightly in different processes. The variations in product yields are primarily because of differences in coal rank and type procesed, one- or two-stage liquefaction concept employed, use of catalyst, operating severities chosen and due to inherent peculiarities of the processes. Also, the quality of product are different.

In one-stage processes (DOW and CO-Steam), in addition to liquid products, a substantial amount of hydrocarbon gases are also produced. Hydrogen consumption in these processes is high, mainly due to higher yields of hydrocarbon gases.

In two-stage processes, the overall picture of product distribution and hydrogen consumption is, however, different. Here the coal is first transformed into a deashed product which is then catalytically hydrocracked to produce clean, light distillate fuel.

A different approach has been used to carry out coal hydrogenation in the first stage, particularly in the CSF, SRC and TSL processes.

The CSF process employs a low-pressure coal extraction concept (first-stage) followed by an ebullating-bed hydrocracking of coal extract. That is, in the first stage, coal slurried in a donor solvent is extracted in a non-hydrogen atmosphere. At a residence time of 40 minutes, about 85 wt% MAF coal is extracted which, after second-stage hydrocracking, yields 33 wt% naphtha (89 RON clear gasoline) and 13.7 wt% hydrocarbon gases on MAF coal basis. The total hydrogen consumption is about 6 wt% MAF coal.

In the SRC process, first-stage involves thermal dissolution of coal in the presence of molecular hydrogen at high temperature and pressure. As a result, a deashed heavy liquid called SRC (which is solid at room temperature) is produced which is then subjected to catalytic hydrocracking in the second stage. At selected liquefaction conditions and 30 minutes residence time, about 60 wt% (MAF basis) of SRC is produced with 2.7 wt% hydrogen consumption. After the second-stage, hydrocracking of SRC, the overall gaseous hydrocarbons and liquid product yields are 6 wt% and 54 wt% MAF coal respectively. The overall hydrogen consumption is about 5.8 wt% MAF coal.

The SRL process is similar to SRC process for both stages except that the feedstock is lignite and part of the hydrogen required is produced in-situ from syngas. The overall yields are 19 wt% hydrocarbon gases and 48 wt% total liquid based on MAF coal. Second-stage hydrogen consumption data are not available.

In the Lummus TSL process, coal dissolution and SRC formation are carried out under similar conditions as in the SRC process with the exception that dissolution takes place in the preheating step without the use of a separate dissolver (reactor) at contact times as short as three to six minutes. The short contact time thermal dissolution concept reduces yields of C_1-C_4 hydrocarbon gases, thus, less hydrogen is consumed in the first stage.

As a result, hydrogen savings from the first stage permit more hydrogen to be utilized into the production of desirable liquid distillates in the second stage. In other words, the overall hydrogen consumption is managed more efficiently. First stage hydrogenation under six minutes residence time yields only 3 wt% (MAF basis) hydrocarbon gases. The hydrogen consumption is significantly low (0.3 wt% MAF coal). The overall total liquid yields, as shown in Table 5-1, are 69 wt% MAF coal with the total hydrogen consumption of 5.1 wt% MAF coal.

In interpreting the hydrogen consumption data, Lummus suggests that efficient hydrogen management in the liquefaction process may be better understood by considering 'hydrogen efficiency' defined as kilograms of total liquid products produced per kilogram of hydrogen consumed. From the yield data in

Table 5-1, the 'hydrogen efficiency' is 6.3, 5.6, 9.3, 10.3 and 13.5 for Z-H, CSF, SRC, DOW and TSL processes respectively indicating that TSL process has certainly the most efficient hydrogen management whereas the CSF has the least. However, such data must be interpreted with caution because the quality of liquid products produced in the processes with high hydrogen efficiency may not necessarily be the best quality. This may be evident from the data given in Table 5-3 where the quality of whole liquid produced is best for CSF and the low for TSL from H/C atomic ratio standpoint (1.73 in CSF versus 1.36 in TSL). It is true that the quality of liquid should also be judged from the heteroatom content (oxygen, sulfur and nitrogen). However, due to insufficient data, such a comparison is not possible.

From Table 5-2, the yields of pyrolysis liquid (tar plus oil) from Occidental process is 27 wt% MAF coal along with 48 wt% of char. The pyrolysis liquids are neither stable nor suitable final products and must be hydrotreated to yield various distillates. Yields reported in Table 5-2 are typical yields after hydrotreatment. Pyrolysis liquid product yields, however, must not be compared with those from liquefaction processes because pyrolysis produces char as the major product.

The yield data for 'coal-oil' processes is also shown in Table 5-2. The Japanese solvolysis process (JSCL) is based on extraction concept where hydrogen is transferred through a solvent. About 62 wt% (MAF basis) of coal extract is produced in the first stage. This extract (which is similar to SRC) is solid at ambient temperature and requires hydrotreating to produce distillate oil fractions. The hydrotreating yield data are not available.

5.0 SUMMARY - CONCLUSIONS AND RECOMMENDATIONS (cont'd)

For the COIL process, yields are 15 wt% naphtha and 59 wt% fuel oil based on MAF coal.

The hydrogen consumption for JSCL and COIL processes is not available.

- Conclusions and Recommendations

All the processes and concepts reviewed in this volume (except SRC-1) were considered for commercializaation not earlier than around 1995. The list of processes and concepts, their R & D status and demonstration scale are summarized in Table 5-3*. Also, the status of funding for further development work to be carried out by corresponding process developers is presented in this table.

It may be seen, that for most of the processes considered in this volume, the process development work is discontinued (Co-Steam, CSF, Zinc Halide, SRL, COIL, Occidental). Financial support for DOW and TSL processes is defined till the end of 1982 only. CE-Lummus, the process developer of TSL process, is negotiating with the U.S. Department of Energy (DOE) for futher support. For Shell and U-Gas gasifiers, the funds for further work toward commercialization are not defined, although research programs are in progress.

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* The Westinghouse gasifier was excluded from this summary and reviewed in the Volume III of the Study report together with other gasifiers considered for commercialization by 1990, due to the accelerated development program initiated at the end of 1981 - beginning 1982 by the process developer, Westinghouse Electrical Corporation, U.S.A., including installation of 1200 t/d gasifier at SASOL II complex scheduled for 1985.

5.0 SUMMARY - CONCLUSIONS AND RECOMMENDATIONS (cont'd)

The cutbacks in financial support are mostly due to the temporary oil glut and economic difficulties in the world, in the U.S.A. particularly.

Although at present the funding for 5500 t/d SRC-1 demonstration plant has been committed by DOE only till the end of 1982, the support likely will be extended in order to complete this demonstration project. Governmental support for further development of Solvolysis process has not been finalized yet. However, according to the national energy policy of Japan, funding for synthetic fuels development should be available.

Among the processes and concepts reviewed, only two have been investigated and tested with lignites (CO-Steam and SRL). However, some conclusions and recommendations for future work on liquefaction of Saskatchewan lignites can be made on the basis of this state-of-the-art review.

- The two-step approach considered in five of the processes and concepts (CSF, TSL, SRC, Zinc Halide and SRL) allows to carry out both the first step (coal liquefaction followed by separation of mineral matter and unreacted coal) and the second step (catalytic hydrocracking of the mineral-free product) at optimum operating conditions of temperature, pressure, residence time, etc. to achieve desirable liquid product distribution, yields and quality with minimum gaseous hydrocarbons produced.
- Catalyst deactivation during hydrocracking of mineral-free product is as low as in a conventional refinery.

Table 5-3: Process Demonstration Scale and Status of New Processes or Concepts Considered for Commercialization around 1995

Process	Demonstration Scale	R&D Status	Funding Status
Direct Coal Liquefaction (One-Stage Hydrogenation)			
- Dow	0.1 t/d	In progress	Current (till Dec. 82)
- CO-Steam	2.3 kg/h (Dismantled)	Abandoned	None
Direct Coal Liquefaction (Two-Stage Hydrogenation)			
- Consol Synthetic Fuel (CSF)	9.0 t/d (Dismantled)	Abandoned	None
- Two-Stage Liquefaction (TSL)	0.5 t/d	In Progress	Current (till Dec. 82)
- Solvent-Refined Coal (SRC)	45 t/d	Demo. Plant Eng. Design In Progress	Current (till Dec. 82)
- Zinc Halide (Z-H)	1.0 t/d	Abandoned	None
- Solvent-Refined Lignite (SRL)	0.5 t/d Dismantled	Abandoned	None
Indirect Coal Liquefaction (Gasification Processes)			
- Shell	150 t/d	In Progress	Not Determined
- U-Gas	25 t/d	In Progress	Not Determined
Coal-Oil Coprocessing			
- Coil	30 bbl/d	Abandoned	None
- Japanese Solvolysis Coal Liquefaction (JSCL)	1.0 t/d	In Progress	Not Determined
- Veba Coal/Oil Process	auto-clave	In Progress	Not Determined
Pyrolysis			
- Occidental	2.5 t/d	Abandoned	None

5.0 SUMMARY - CONCLUSIONS AND RECOMMENDATIONS (cont'd)

- In a two-step coal liquefaction process, the hydrogen efficiency (defined as kilogram of total liquid product produced per kilogram of hydrogen consumed), may be quite high depending on liquid product slate and quality required (e.g. for SRC-1 producing estimated 44 wt% naphtha, 44 wt% mid-distillate and 12 wt% heavy oil, the hydrogen efficiency is 9.3. For TSL producing estimated 4 wt% naphtha, 12 wt% light oil; 61 wt% mid-distillate and 23 wt% heavy oil, the hydrogen efficiency is 13.5.

- Co-Steam and SRL process development may be considered as a supporting research for processing of low-rank coals and lignites for other more advanced coal liquefaction processes being developed for higher rank coals (e.g. SRC-1).

- The DOW concept of using emulsified catalyst for coal hydroliquefaction is an attractive feature to be explored in the future work on Saskatchewan lignite liquefaction.

- Occidental pyrolysis process has a number of advantages in comparison to other pyrolysis processes e.g. lighter liquid product due to the rapid quenching of product vapors by direct contact with a process derived hydrogen-donor solvent ("polystop" quench solvent), fine granulated char, etc. This process concept may be recommended for futher investigation for Estevan lignite if a limited total liquid product yields (about 20 wt% MF coal) is acceptable and char can be marketed and/or utilized effectively in a power plant.

- Co-processing option of coal liquefaction has a number of potential merits, especialy for unique situation in the Province of Saskatchewan where both the lignite and heavy-oil resources are available. However, till now, the co-processing of coal and heavy oil has been investigated mostly at autoclave/bench scale level only. Further, R & D including PDU and pilot-plant scale is recommended.

- For indirect coal liquefaction, Shell entrained-flow, pressurized gasifier has a number of attractive features for production of syngas from Saskatchewan lignites e.g. elevated operating pressure, negligible amount of methane in the raw gas, no restrictions on fines in the feed coal, ability to gasify relatively high ash content coal, etc. However, till now, lignites have not been tested in this gasifier. The process is not as advanced as other gasification processes and due to the recent cutbacks, the readiness of this process towards commercialization may be delayed even beyond 1995.

REFERENCES AND SOURCES USED

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REFERENCES AND SOURCES

General

H.I. Smith, "Coal Conversion Technology", Chemical Technology Review No. 66, Noyes Data Corporation, N.J., 1976.

J.D. Batchelor, U.S. Department of Energy, "Technical Status of the Large Coal Liquefaction Programs at the Department of Energy", paper presented at Institute of Gas Technology sponsored Symposium on Advances of Coal Utilization Technology, Louisville, Kentucky, May 14-18, 1979.

N. Berkowitz, "Introduction to Coal Technology", Academic Press, N.Y., 1979.

P. Nowacki, "Coal Liquefaction Processes", Chemical Technology Review No. 131, Energy Technology Review No. 45, published by Noyes Data Corp. N.J., 1979.

P. Sterbenz, "Coal Conversion Today", Hydrocarbon Processing, p.68-72, July 1980.

B.R. Rodger et al., Oakridge National Laboratory, "Assessment of Current Research and Development in Support of the U.S. Coal Liquefaction Demonstration Plants Program", paper presented at the 15th Intersociety Energy Conversion Engineering Conference, Seattle, Washington, August 18-22, 1980. Proceedings Volume 3, p. 1808-1811.

"Low Rank Coal Study, National Needs for Resource Development, Vol. 3 - Technology Evaluation". Report prepared for U.S. Department of Energy under the Contract No. DE-AC18-79 FC 10066, Nov. 1980.

D. Olliver, "Oil and Gas from Coal", published by Financial Times Business Information Ltd., 10 Cannon Street, London, Jan. 81.

"Coal Liquefaction - The Chemistry and Technology of Thermal Processes", Academic Press 1980.

DOW Process

G.J. Quarderer and N.G. Moll, The Dow Chemical Co., "Coal Hydrogenation Catalyst Recycle". U.S. Patent 4, 090, May 23, 1978.

G.J. Quardere and N.G. Moll, The Dow Chemical Co., "Conversion Process for Solid Hydrocarbonaceous Materials". U.S. Patent 4, 102, 775, July 25, 1978.

"Dow Unveils Coal Liquefaction Process", Chemical Week, September 13, 1978.

DOW Process (cont'd)

"Dow Looks to Scale Up Coal Liquefaction Process", *Chemical Week*, September 20, 1978.

N.G. Moll and G.J. Quarderer "The Dow Coal Liquefaction Process", Chemical Engineering Progress, pp. 46-50, November 1979.

G.J. Quarderer and N.G. Moll, The Dow Chemical Co., "Emulsion Catalyst for Hydrogenation Process". U.S. Patent 4, 136, 013, June 23, 1979.

G.J. Quarderer and N.G. Moll, The Dow Chemical Co., "Process for Recovering a Premium Oil from a Slurry Produced by High Temperature Hydrogenation of a Solid Hydrocarbonaceous Fuel" U.S. Patent 4, 180, 456, Dec. 25, 1979.

G.J. Quarderer and N.G. Moll, "Separation of Solids from Coal Liquefaction Products via Counter-Current Solvent Desaphlting", *Separation Science and Technology*, *15* (3), pp. 165-183, 1980.

G.A. Stuart, "Economics of the Dow Coal Liquefaction Process", presented at 49th AIChE meeting, Philadelphia, May 1980.

G.A. Stuart, Dow Chemical Company, Midland, Michigan, Personal Correspondence Sept. 19, 1980.

N.G. Moll et al., "Dow Coal Liquefaction Process Update", synthetic Fuels Status and Direction Conference, San Francisco, CA, October 13-16, 1980.

CO-Steam Process

H.R. Appell, and I. Wender, "Hydrogenation of Coal with Carbon Monoxide and Water", American Chemical Society, Div. of Fuel Chemistry Preprints, Vol. 12, No. 3, pp. 220-222, 1968.

E. Del Bel, S. Friedman, P.M. Yavorsky, and I. Wender. The Liquefaction of Lignite by the CO-Steam Process. AlChE National Meeting, Houston, Texas, March 16-20, 1975.

E.A. Sondreal, C.L. Knudson, J.E. Schiller, and T.H. May. Development of the CO-Steam Process for Liquefaction of Lignite and Western Subbituminous Coals, Proceedings, 1977 Symposium on Technology and Use of Lignite, Grand Forks, ND, May 1977.

D.E. Severson, A.M. Souby and J.J. Harris. Process Development of Lignite Liquefaction, Proceedings 1977 Symposium on Technology and Use of Lignite, Grand Forks, ND, May 1977.

DOW Process (cont'd)

"Dow Looks to Scale Up Coal Liquefaction Process", *Chemical Week*, September 20, 1978.

N.G. Moll and G.J. Quarderer "The Dow Coal Liquefaction Process", Chemical Engineering Progress, pp. 46-50, November 1979.

G.J. Quarderer and N.G. Moll, The Dow Chemical Co., "Emulsion Catalyst for Hydrogenation Process". U.S. Patent 4, 136, 013, June 23, 1979.

G.J. Quarderer and N.G. Moll, The Dow Chemical Co., "Process for Recovering a Premium Oil from a Slurry Produced by High Temperature Hydrogenation of a Solid Hydrocarbonaceous Fuel" U.S. Patent 4, 180, 456, Dec. 25, 1979.

G.J. Quarderer and N.G. Moll, "Separation of Solids from Coal Liquefaction Products via Counter-Current Solvent Desaphlting", *Separation Science and Technology, 15* (3), pp. 165-183, 1980.

G.A. Stuart, "Economics of the Dow Coal Liquefaction Process", presented at 49th AIChE meeting, Philadelphia, May 1980.

G.A. Stuart, Dow Chemical Company, Midland, Michigan, Personal Correspondence Sept. 19, 1980.

N.G. Moll et al., "Dow Coal Liquefaction Process Update", synthetic Fuels Status and Direction Conference, San Francisco, CA, October 13-16, 1980.

CO-Steam Process

H.R. Appell, and I. Wender, "Hydrogenation of Coal with Carbon Monoxide and Water", American Chemical Society, Div. of Fuel Chemistry Preprints, Vol. 12, No. 3, pp. 220-222, 1968.

E. Del Bel, S. Friedman, P.M. Yavorsky, and I. Wender. The Liquefaction of Lignite by the CO-Steam Process. AIChE National Meeting, Houston, Texas, March 16-20, 1975.

E.A. Sondreal, C.L. Knudson, J.E. Schiller, and T.H. May. Development of the CO-Steam Process for Liquefactio of Lignite and Western Subbituminous Coals, Proceedings, 1977 Symposium on Technology and Use of Lignite, Grand Forks, ND, May 1977.

D.E. Severson, A.M. Souby and J.J. Harris. Process Development of Lignite Liquefaction, Proceedings 1977 Symposium on Technology and Use of Lignite, Grand Forks, ND, May 1977.

CO-Steam Process (cont'd)

W.G. Wilson, et al., "Application of Liquefaction Processes to Low-Rank Coals", Ind. & Eng. Chem. Process Research and Development, Dec. 1979.

W.G. Willson, et al., "Comparison of Liquefaction Yields", Proceedings, 1981 Symposium on Technology & Use of Lignite, San Antonio, Texas, June 1981.

CSF Process

Fluor Engineers and Contractors, Inc., "Conceptual Design for Advanced Coal Liquefaction Commerical Plant", Final Report prepared for U.S. Department of Energy under Contract No. EX-76-C-01-2251, Report No. FE-2251-40 (Vol. 1), Oct. 1977.

J.A. Kleinpeter et al., "Coal Liquefaction and Deashing Studies of the CSF and SRC Processes", Report prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1517, Report No. FE-1517-T68, Nov. 1978.

Two-Stage Liquefaction (TSL) Process

M.C. Sze and G.J. Snell, The Lummus Company, "Coal Liquefaction", U.S. Patent 3,852,182, Dec. 3, 1974.

M.C. Sze and G.J. Snell, The Lummus Company, "Coal Liquefaction", U.S. Patent 3,856,675, Dec. 24, 1979.

M.B. Neuworth and E.C. Moroni, "Development of a Two-Stage Liquefaction Process", Advances in Coal Liquefaction Technology Symposium sponsored by the Institute of Gas Technology, Louisville, Kentucky, May 14-18, 1979.

E.M. Phillips et al., "A Comparative Study of Coal Liquefaction Performance and Economics for Solvent Refined Coal-Based Processes", 88th National Meeting, American Institute of Chemical Engineers - Fuels and Petrochemical Division, Philadelphia, June 8-12, 1980.

H.D. Schindler et al., "Advanced Coal Liquefaction Processes Emphasize Low Hydrogen Consumption" 15th Intersociety Energy Conversion Engineering Conference, Seattle, Washington, pp. 1543-1556, Aug. 18-22, 1980.

H.D. Schindler et al., "The Integrated Two-Stage Liquefaction Process", AIChE Annual Meeting, New Orleans, Louisiana, Nov. 1981.

J.D. Potts, et al., "LC-Fining of Solvent Refined Coal", ACS Symposium Series 156 - Upgrading Coal Liquids, Editor R.F. Sullivan, American Chemical Society, 1981.

Two-Stage Liquefaction (TSL) Process (Cont'd)

"Technical Data - 30,000 T/D Integrated Two Stage Coal Liquefaction Plant", C.E. Lummus Company, Bloomfield, N.J., Jan. 1982.

Solvent Refined Coal (SRC) Process

B.K. Schmid, "Status on SRC Project", Chemical Engineering Progress, Vol. 71, No. 4, pp. 75-78, April 1975.

G.E. Chenoweth, "History of the SRC Pilot Plant", 2nd Annual Conference on Coal Gasification and Liquefaction, U. of Pittsburg, Aug. 5-7, 1975.

G.R. Pastor, et al., "Operation of the SRC Pilot Plant", Chemical Engineering Progress, pp. 143-144, Aug. 1976.

R.M. Adams, et al., "Critical Solvent Deashing of Liquefied Coal", Chemical Engineering Progress, pp. 44-48, June 1979.

M. Peluso and D.F. Ogren, "Antisolvent Deashing", Chemical Engineering Progess, pp. 41-43, June 1979.

M. Pell, et al., 14th Intersociety Energy Conversion Conference, Boston, Aug. 5-10, 1979.

B.K. Schmid and D.M. Jackson, "Liquid and Solid Fuels by Recycle SRC", Coal Processing Technology, Vol. 5, pp. 146-155, 1979.

R.C. Chillingworth, K.E. Hastings and J.D. Potts, "Alternative Modes of Processing SRC in an Expanded Bed LC-Finer to Produce Low Nitrogen Distillates", Ind. & Eng. Chem. Product Research and Development, Vol. 19, pp. 31, March 1980.

J.D. Potts, H. Unger and E.M. Phillips, "Application of LC-Fining Technology for Upgrading Coal Extracts in a Two-Stage Liquefaction Concept", 179th National ACS Meeting, Division of Fuel Chemistry, Houston, Texas, March 23-28, 1980.

J.C. Tao, A.D. Bixler and E.J. Breyne, " Demonstration and Commercialization of the SRC-I Technology", Seventh Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity, Pittsburgh, PA., August 5, 1980

H.M. Risbud, "Technical Comparison Among Solids Separation Alternatives in SRC-I Process", EPRI Contractor's Conference, Palo Alto, CA, May 14, 1981.

J.R. O'Leary and G.C. Rappe, "Scale-up of an SRC Deashing Process", Chemical Engineering Progress, pp. 67-72, May, 1981.

Solvent Refined Coal (SRC) Process (cont'd.)

S.V. Raman, G.C. Rappe, and S.C. Weiner, "Research and Development Support for a Solvent Refined Coal Demonstration Plant - An Overview", Second World Congress of Chemical Engineering, Montreal, Canada, Oct. 8, 1981.

R.M. Thorogood, C.L. Yeh, S.M. Morris, "Scaleup of the SRC-1 Coal Liquefaction Process", Second World Congress of Chemical Engineering, Montreal, Canada, Oct. 8, 1981.

Zinc-Halide (Z-H) Process

Conoco Coal Development Company, "Zinc Halide Hydrocracking Process for Distillate Fuels from Coal", Quarterly Technical Progress Report, Feb. 1 - April 30, 1978, Report prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1743, Report No. FE-1743-53, May 20, 1978.

Conoco Coal Development Company, "Zinc Halide Hydrocracking Process for Distillate Fuels from Coal", Quarterly Technical Progress Report, May 1 - July 31, 1978, Report prepared for U.S. Department of Energy under Contract No. EX-76-C-01-1743, Report No. FE-1743-57, Aug. 20, 1978.

Conoco Coal Development Company, "Zinc Halide Hydrocracking Process for Distillate Fuels from Coal", Quarterly Technical Progress Report, Aug. 1 - Oct. 31, 1978, Report prepared for U.S. Department of Energy, under the Contract No. EX-76-C-01-1743, Report No. FE-1743-62, Nov. 20, 1978.

Conoco Coal Development Company, "Design and Construction of the PDU for Zinc Halide Hydrocracking of Coal, Final Report for Tasks 4 and 5", Report prepared for U.S. Department of Energy Under Contract No. EX-76-C-01-1743, Report No. FE-1743-61, Dec. 8, 1978.

R. Salmon, et al., "Technical and Economic Evaluation of Selected Coal Liquefaction Processes. Phase 0: Preliminary Screening Evaluations", Oak Ridge National Laboratory Report ORNL/TM-7581, April 1981.

R.T. Struck and C.W. Zidke, "Hydrocracking of Coal to Light Distillate with Molten Zinc Chloride", Fuel, Vol. 60, Sept. 1981.

Solvent Refined Lignite (SRL) Process

K.J. Klabunde, et al., "Production of Clean Liquid Fuels from Lignite: Catalytic Hydrotreating of Solvent Refined Lignites Employing Low Surface Area Catalysis and Commercial Cataysts", Symposium on Technology and Use of Lignite, Grand Forks, N.D., May 1977, GFERC/IC-77/1, pp. 193-206.

Solvent Refined Lignite (SRL) Process (cont'd)

D.E. Severson, A.M. Souby and J.J. Harris, "Process Development of Lignite Liquefaction", Symposium on Technology and use of Lignite, Grand Forks, N.D., May 1977, GFERC/IC-77/1, pp. 159-192.

Occidental Pyrolysis Process

E.W. Everette, et al., "Flash Pyrolysis Coal Liquefaction Process Development - Quarterly Report April - June 1977", Report prepared for U.S. Department of Energy under the Contract No. EX-76-C-01-2244, Report No. FE-2244-14.

K. Durai-Swamy, and W. Deslate, "Occidental's Flash Pyrolysis Coal Liquefaction Process and It's Application to Power Generation", EPRI Conference on Coal Pyrolysis, Palo Alto, California, Feb. 25-26, 1981.

Z. Kolisnyk and R.A. Grant, "Coal Pyrolysis - The Third Alternative", 33rd Canadian Conference on Coal, Regina, Saskatchewan, Sept. 20-22, 1981.

COIL Process

G.R. De Vaux, E.S. Johanson, and M.C. Chervenak, "COIL Process-First Generation Commercial Coal Liquefaction Plant", Miami International Conference on Alternative Energy Sources, Vol. 7, pp. 3267-3283, 1977.

U.S. Patent 3,755,137, "Multi-Stage Ebullated Bed Coal-Oil Hydrogenation and Hydrocracking Process", Aug. 28, 1973, Assigned to Hydrocarbon Research Inc., New York, N.Y.

Japanese Solvolysis (JSCL) Liquefaction Process

S. Arita, K. Osafune, Y. Yamada, H. Kakiyama, H. Houda, N. Tagawa, and S. Abe, S. "Solvolysis Liquefaction of Coal Using Petroleum vacuum residue: II Effect of Rank of Coal on Yield". Reports of the Govt. Ind'l Research Inst., Kyushu, No. 19, Nov. 1977.

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APPENDIX A

ULTIMATE, PROXIMATE, AND PETROGRAPHIC ANALYSIS OF DIFFERENT RANK COALS

TABLE A-1 Ultimate, Proximate, and Petrographic Analysis of Different Rank Coals

COAL	LIGNITE				
	Saskatchewan		North Dakota		Texas
	'Estevan'	'Willow Bunch'	'Indian Head'	'Beulah'	'Big Brown'
Moisture (As-Received), wt%	35.0	35.0	33.6	28.8	28.4
Heating Value (GCV) MF Coal, KJ/kg	24272	20380	21149	20163	22093
Proximate Analysis, wt% MF Coal					
Volatile Matter	42.2	40.0	44.1	40.6	44.4
Fixed Carbon	42.2	38.8	45.7	43.0	39.2
Ash	15.6	21.2	10.2	16.7	16.4
Ultimate Analysis, wt% MAF Coal					
Carbon	72.3	66.0	71.1	69.6	74.2
Hydrogen	5.3	5.1	5.2	4.4	5.7
Nitrogen	1.2	0.6	1.0	1.0	1.3
Sulfur	0.7	1.1	1.3	2.8	1.4
Oxygen	20.5	27.2	21.4	22.2	17.4
Petrographic Analysis, vol%					
Vitrinite	80	N.A.	80.5	N.A	65.3
Exinite	10		2.7		16.3
Inertinite	10		16.8		18.4
Ash Analysis, wt%					
Silica, SiO_2	36.6	42.1	26.3	27.6	40.6
Alumina, Al_2O_3	21.8	15.6	10.7	14.0	14.4
Titania, TiO_2	1.3	0.9	0.5	0.6	1.2
Ferric Oxide, Fe_2O_3	4.8	6.7	14.0	14.2	7.8
Lime, CaO	15.1	14.3	16.0	14.5	13.2
Magnesia, MgO	3.7	5.0	4.9	3.9	3.7
Potassium Oxide, K_2O	0.4	1.3	0.9	0.4	0.7
Sodium Oxide, Na_2O	6.5	1.8	8.0	5.5	0.5
Phosphorous Pentoxide, P_2O_5	0.5	–	18.4	0.2	0.6
Sulfur Trioxide, SO_3	9.3	12.3	–	19.1	16.8
Undetermined					

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TABLE A-1 Ultimate, Proximate, and Petrographic Analysis of Different Rank Coals (cont'd)

COAL	SUBBITUMINOUS		BITUMINOUS			
	Alberta 'Forestburg'	Wyoming 'Wyodak'	Illinois 'Illinois #6'	Pennsylvania 'Pittsburg #8	Kentucky 'Kentucky #9'	Indiana 'Indiana V'
Moisture (As-Received) wt%	25.2	30.0	11.0	1.0	10.0	
Heating Value (GCV) MF Coal, KJ/kg	27530	26563	26763	32634	29657	
Proximate Analysis, wt% MF Coal						
Volatile Matter	41.2	44.6	37.9	42.7	39.8	N.A.
Fixed Carbon	53.6	46.8	51.5	47.4	49.8	N.A.
Ash	5.2	8.6	10.6	9.9	10.4	10.6
Ultimate Analysis, wt% MAF Coal						
Carbon	74.4	73.1	79.3	80.6	78.8	78.1
Hydrogen	4.6	5.5	5.5	5.7	5.6	5.3
Nitrogen	1.4	0.9	1.7	1.2	1.6	1.6
Sulfur	0.5	1.1	3.9	5.4	3.8	3.8
Oxygen	19.1	19.4	9.6	7.1	10.2	11.2
Petrographic Analysis, vol%						
Vitrinite	71.4	N.A.	N.A.	N.A.	N.A.	
Exinite	7.3					
Inertinite	21.3					
Ash Analysis, wt%						
Silica, SiO_2	19.7	29.3	47.8	N.A.	N.A.	N.A.
Alumina, Al_2O_3	14.6	18.8	18.1			
Titania, TiO_2	8.5	1.1	0.8			
Ferric Oxide, Fe_2O_3	30.3	6.6	17.8			
Lime, CaO	4.5	18.8	7.4			
Magnesia, MgO	1.0	4.5	0.9			
Potassium Oxide, K_2O	0.9	0.4	1.8			
Sodium Oxide, Na_2O	0.2	0.6	0.5			
Phosphorous Pentoxide, P_2O_5	0.4	0.4	–			
Sulfur Trioxide, SO_3	18.4	17.3	3.7			
Undetermined		1.7	1.2			

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APPENDIX B

CORRESPONDENCE WITH PROCESS DEVELOPERS

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Montreal, Canada H5B 1C8

Dear Josef:

As part of your engineering study on Saskatchewan coal utilization, you have asked for information on our COIL™ process. Although it is not the purpose of this letter to provide a full discourse on the subject, I hope to answer a few of the questions which you expressed in our recent telephone conversation. I will assume that you are familiar with the basic description of the process, based on previous conversations at HRI, as well as literature which we have left with you on other occasions.

Unfortunately, we have no bench unit or PDU data on which to draw, as we have with H-Coal®. Therefore, we cannot offer to predict, or even approximate, COIL results from microautoclave tests, as we have offered to do in the case of H-Coal. It is our belief that no meaningful conclusions could be drawn from data obtained in microautoclave solutions of Saskatchewan coal and oil. However, if you wish to draw your own conclusions from raw data, we will run the experiment for you. We have discussed the possibility of using COIL with other clients before, and have always come to the conclusion that at least one bench scale experiment will be needed before we can discuss the merits of this process relative to straight H-Coal or H-Oil™ processing. Even with one bench scale run, we would not be in a position to develop a formal process flowsheet. This type of process development will take many months, and a high level of laboratory evaluation and process development work. If

Telex: 83-1731 Cable: HYDROCOL A subsidiary of DYNALECTRON CORPORATION

SNC is interested in further evaluation of our COIL process, we would be happy to meet with you and discuss arrangements for such.

It has been our experience with other clients that economics, more than technical specifications, determine the relative value of the COIL process. When comparing the COIL process with H-Coal for a given set of feedstocks (coal and oil), it will probably be a matter of the relative economic value of the feedstocks and products, processed independently versus processed simultaneously, which will determine the relative value of each process. Again, I am answering you with broad generalties, but because of our very limited data base, I cannot be more specific.

If you wish to supply us with a nonconfidential description of the coal and oil which you have available, our technical staff in the Liquid Phase Hydrogenation Department will be happy to provide an educated guess about the viability of their use as COIL feedstocks. If you wish to discuss this subject further, please let me know how HRI may be of service.

Sincerely,

Jeff Gendler

Jeffrey L. Gendler
HRI R&D Marketing

JLG/jah

SASKOIL COAL LIQUEFACTION STUDY

Review and Evaluation of Indirect Liquefaction Processes with the Following Gasifiers:

Westinghouse
U-Gas
Shell-Koppers

December 1981

Foster Wheeler Synfuels Corporation
110 SOUTH ORANGE AVENUE, LIVINGSTON, NEW JERSEY

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

SASKATCHEWAN COAL LIQUEFACTION STUDY

SECOND GENERATION GASIFIERS WESTINGHOUSE, U-GAS AND SHELL-KOPPERS

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C. Process Descriptions

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D. Comparative Investment Costs

E. Comments

F. Tabulation of Significant Results

G. References

INTRODUCTION

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

INTRODUCTION

This report deals with the comparison of various indirect coal liquefaction processes which have their gasification performed by second generation gasifiers.

The gasifiers here considered are: Westinghouse, U-Gas and Shell-Koppers.

The ultimate purpose of the processes comparison is the selection of the most suitable one for the specific coal under consideration and the geographical area for the facility.

The basic assumption for the prescreening study are:

1. Coal: Lignite from the Estevan zone and Willow Bunch coal fields with a proximate and ultimate analysis as shown in other parts of this report.

   Coal feed to the Plant: 20,000 metric tons per stream day, as received.

2. Coal will be available at the Facility Battery Limits, being shipped from the mine by rail, truck or conveyors.

3. The Facility is a grass roots one and only coal, water and air are required as major raw materials.

   It is assumed that the water requirements will be met from wells or otherwise.

4. Ash/slag will be disposed of on-site or off-site. Sufficient acreage will be made available for 20 years storage, if on-site disposal.

5. Product Quality will be as produced within the Battery Limits, with no upgrading or secondary refining.
6. Electrical Power in excess of the captive requirements will be available for export.
7. Plant on stream factor: 90%.

The following cases are considered in this report:

Case A: Westinghouse Gasifier with Fischer-Tropsch Synthesis.

Case A-R: Maximization of liquid fuels by reforming the SNG produced in Case A.

Case B: U-Gas Gasifier with Fischer-Tropsch Synthesis.

Case B-R: Maximization of liquid fuels by reforming the SNG in Case B.

Case C: Shell-Koppers Gasifiers with Fischer-Tropsch Synthesis.

Case C-R: Maximization of liquid fuels by reforming the SNG in Case C.

Case C-1: Shell-Koppers gasifier with methanol synthesis followed by MRDC* methanol to gasoline process.

---

* Mobil Research and Development Corporation.

A Saskatchewan lignite from the Estevan zone with the following proximate and ultimate analysis was used in this report:

Proximate Analysis

	Weight %
Volatile Matter	27.30
Fixed Carbon	27.30
Moisture	35.00
Ash	10.01
Sulfur	0.39

Ultimate Analysis

	Weight %
H	2.60
C	39.78
N	0.65
O	11.57
S	0.39
Ash	10.01
H_2O	35.00

A lignite from the Willow Bunch Coal Field with the following proximate and ultimate analysis was also considered in this report:

Proximate Analysis

	Weight %
Volatile Matter	27.54
Fixed Carbon	35.03
Moisture	35.00
Ash	13.65
Sulfur	0.58

Ultimate Analysis

	Weight %
H	2.61
C	33.86
N	0.32
O	13.98
S	0.58
Ash	13.65
H_2O	35.00

This coal has a higher ash and oxygen and lower carbon content than the Estevan coal.

This will result in lower yields of products than the Estevan lignite.

No attempt was made to estimate the actual yields, thermal efficiencies and other parameters for the Willon Bunch coal.

In general terms, it may be said that the expected products yields from the Willow Bunch would be about 85% of the ones for the Estevan lignite.

EXPECTED PRODUCTS QUALITY

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

EXPECTED PRODUCTS QUALITY

GASOLINE FROM METHANOL BY THE MRDC PROCESS

COMPARISON OF PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS WITH ESTIMATED METHANOL-TO-GASOLINE GASOLINE PROPERTIES-

	Estimated MTG Unleaded Gasoline Properties	Specification
Gravity, °API	61.4	
Octane Numbers		
Research	93	
Motor	83	82 min.
(Research + Motor)/2	88	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation, °F		
IBP	85.5	
10%	110.5	158 max.
30%	146.1	
50%	200.0	170/250 min./max
70%	262.8	
90%	336.9	374 max.
EP	388.6	437 max.
V/L Ratio (=20), °F	@ 129.5	@ 140 max.
Sulfur, wt. %	Nil	0.10 max.
Composition, vol. %		
Paraffins	51.3	
Olefins	11.3	20 max. (target)
Naphthenes	8.7	
Aromatics	28.7	
Durene Content, wt. %	4.6	5 max. (target)
Molecular Weight	94.3	

SOURCE: D.O.E. FE-2447-9, Jan. 1978

PRODUCTS QUALITY

SNG: BY LURGI (D.B.) TO GASOLINE VIA METHANOL

Composition

Hydrogen	1.7
Methane	95.5
Ethane	0.2
Propane	0.1
Butane	0.1
Carbon dioxide	0.5
Inerts (N_2 + Ar)	1.9
	100.0

Other

Heat of combustion (HHV)	980 Btu/SCF
Carbon monoxide (0.1% max.)	0.02%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

Index	Calculated	Preferable	Objectionable
Lifting, I_l	1.03	under 1.0	above 1.06
Flash-back, I_f	1.02	under 1.15	above 1.2
Yellow-tip, I_y	1.05	above 1.0	under 0.8

The SNG product is of satisfactory quality and is compatible with pure methane. The inerts from the oxygen stream for gasification, however, give rise to a lifting index value between the preferable and objectionable values. From a processing standpoint, there is little that can be done to reduce this value. The selection of a more expensive oxygen plant yielding a 99+% pure product, however, would alleviate this problem. (The SNG product, on the other hand, has a lifting index value better than the preferable value when compared with the high methane natural gas standard cited in the A.G.A. Research Bulletin Number 36. This gas has a 5.0% inerts content.)

SOURCE: D.O.E. FE-2447-9, Jan. 1978

Job No.: B-28-8011
Date: November, 1981

PRODUCTS QUALITY

C_3LPG & C_4LPG

COMPARISON OF PRINCIPAL LPG SPECIFICATIONS WITH ESTIMATED METHANOL-TO-GASOLINE LPG PROPERTIES-

	Commercial Propane ASTM D1835*	Est. MTG C_3 LPG Properties	Commercial Butane ASTM D1835*	Est. MTG C_4 LPG Properties
Vapor Pressure at 100°F, max. psig (D1267)	208	171	70	52
Volatile Residue				
Butane and Heavier, max. vol. %	2.5	1.9	-	-
Pentane and Heavier, max. vol. %	-	-	2.0	0.7
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

*1976 Annual Book of ASTM Standards

SOURCE: P.O.E. FE-2447-9, Jan. 1978

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

PRODUCTS QUALITY

GASOLINE FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL UNLEADED GASOLINE SPECIFICATIONS WITH ESTIMATED FISCHER-TROPSCH GASOLINE PROPERTIES -

	Estimated F-T Unleaded Gasoline Properties	Specifications
Gravity, °API	67.2	
Octane Numbers		
Research	91	
Motor	83	82 min.
(Research + Motor)/2	87	87 min.
Volatility		
Reid Vapor Pressure, lb	10.0	
Distillation °F		
IBP	86.5	
10%	108.9	158 max.
30%	137.3	
50%	186.0	170/250 min./max
70%	249.6	
90%	335.2	374 max.
EP	420.6	437 max.
V/L Ratio (=20), °F	@127.3	@140 max.
Sulfur, wt. %	Nil	0.10 max.
Composition, Vol. %		
Paraffins	60.5	
Olefins	19.9	20 max. (target)
Naphthenes	3.1	
Aromatics	16.5	
Molecular Weight	93.1	

SOURCE: D.O.E. FE-2447-9 Jan. 1978

PRODUCTS QUALITY

GASOLINE FROM F-T SYNTHESIS

EXPECTED BLEND COMPOSITION

Blending

Component	Wt. %
Mixed butanes	2.7
Light hydrocarbons	2.9
Alkylate	2.7
Light polymer gasoline	16.3
Hydrogenated heavy polymer gasoline	14.3
C_5/C_6 isomerate	25.1
Reformate	23.8
Hydrotreated gasifier naphtha	11.0
	100.0

Properties

Estimated properties and a comparison with unleaded gasoline specifications are presented elsewhere in this report.

This gasoline meets all specifications. Compared to typical present-day gasolines, it is slightly lower in gravity and contains fewer aromatics.

Heavy polymer gasoline hydrogenation has been added in order to meet the olefin target of 20 vol. % max. Without hydrogenation, the following are estimated:

(Research + Motor)/2 = 88.5
Olefin, vol. % = 28

Mobil and industry experience with high olefinic gasolines is limited. Consequently, marketing such a gasoline would require extensive testing and, perhaps, additive package reformulation.

SOURCE: D.O.E. FE-2447-9, Jan. 1978

PRODUCT QUALITY

DIESEL FUEL FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL DIESEL FUEL SPECIFICATIONS WITH ESTIMATED FISCHER-TROPSCH DIESEL FUEL PROPERTIES

	DIESEL FUEL-ASTM D975[1]			Estimated F-T Diesel Fuel Properties
	Grade No. 1-D	Grade No. 2-D	Other	
Gravity, °API	-	-		56
Cetane Number, min. (D613 or D976)	40	40	45[2]	50
Volatility				
90% min./max. °F (D86)	-/550	540/640[5]		520
Flash Point, °F min. (D93)	100	125		125
Cloud Point, °F (D2500)	(3)	(3)		10-20
Viscosity, Kin. @ 100°F min./max. (D445)	1.4/2.5	2.0/4.3		2
Sulfur, wt. % max. (D129)	0.50	0.50	(4)	None
Corrosion, max. (D130)	No. 3	No. 3		No. 1
Carbon Residue on 10% Resid., max. (D524)	0.15	0.35		Trace

---

(1) 1976 Annual Book of ASTM Standards.
(2) Various engine builders' requirements.
(3) Varies with marketing region; maps included in ASTM D975.
(4) Varies in individual states, some as low as 0.3%.
(5) 540/680 90% min./max. proposed but currently not accepted by industry.

SOURCE: D.O.E., FE-2447-9, January 1978.

Job No.: B-28-8011
Date: November, 1981

FOSTER WHEELER

PRODUCTS QUALITY

SNG EXPECTED COMPOSITION FROM F - T SYNTHESIS

Composition

Hydrogen	3.8
Methane	89.7
Ethene	1.0
Ethane	2.3
Propene	1.0
Propane	0.1
Carbon Dioxide	0.5
Inerts (N_2 + Ar)	1.6
	100.0

Other

Heat of Combustion (HHV)	1,003 Btu/SCF
Carbon monoxide (0.1% max.)	0.07%
Water	0.01%
Sulfur	None

Compatibility Indexes (versus pure methane)

Index	Calculated	Preferable	Objectionable
Lifting, I_1	0.99	under 1.0	above 1.06
Flash-back, I_f	1.06	under 1.15	above 1.2
Yellow-Tip, I_y	0.97	above 1.0	under 0.8

The SNG, product is of satisfactory quality and is interchangable with pure methane. The olefin nature of the Fischer-Tropsch light gases, however, gives rise to a yellow-tip index value slightly below the preferable value. If the olefins prove troublesome in actual tests, an expensive hydrogenation step for olefin saturation could be added.

SOURCE: D.O.E. FW-2447-9, Jan. 1978

PRODUCTS QUALITY

HEAVY FUEL OIL FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL FUEL OIL AND GAS TURBINE FUEL OIL SPECIFICATIONS WITH ESTIMATED FISCHER-TROPSCH HEAVY FUEL OIL PROPERTIES

	Fuel Oil No. 5 (Light) ASTM D396*	Gas Turbine F.D. No. 5 - GT ASTM D2880*	Estimated F-T Hvy. Fuel Oil Properties
Gravity, °API	-	-	41
Volatility			
Flash Point, °F min. (D93)	130 or Legal	130 or Legal	150
Distillation	-	-	-
Ash, wt. % max. (D482)	0.10	0.03	Trace
Viscosity, SSU at 100°F, min. (D445)	150	45	Satisfactory
Sulfur, wt. % (D129)	Legal	-	None
Metals, ppm (wt.) max.			
Vanadium	-	2	0
Sodium + Potassium	-	5	Trace
Calcium	-	10	0
Lead	-	5	0

*1976 Annual Book of ASTM Standards.

SOURCE: D.O.E. FE-2447-9. Jan. 1978

PRODUCTS QUALITY

C_3LPG & C_4LPG FROM F-T SYNTHESIS

COMPARISON OF PRINCIPAL LPG SPECIFICATIONS WITH ESTIMATED FISCHER - TROPSCH LPG PROPERTIES-

	Commercial Propane ASTM D1835*	Est. MTG C_3 LPG Properties	Commercial Butane ASTM D1835*	Est. MTG C_4 LPG Properties
Vapor Pressure at 100°F, max., psig (D1267)	208	178	70	37
Volatile Residue				
Butane and Heavier, max. vol. % (D2163)	2.5	1	-	-
Pentane and Heavier, max. vol. % (D2163)	-	-	2.0	1
Corrosion, max. (D1838)	No. 1	No. 1	No. 1	No. 1
Sulfur, grains (D2784)	15	None	15	None

*1976 Annual Book of ASTM Standards.

Source: D.O.E., FE-24-2447-9, Jan. 1978

PRODUCTS QUALITY

ALCOHOLS EXPECTED COMPOSITION FROM F - T SYNTHESIS

Composition

Composition of the mixed alcohol product is:

Ethanol	59
i-propanol	14
N-propanol	14
2-butanol	4
I-butanol	1
N-butanol	5
2-pentanol	1
N-pentanol	1
C_6 plus	1
	100

The alcohol product is essentially free of acids, aldehydes, ketones and water.

Source, D.O.E. FE-2447-9, Jan. 1978

FOSTER WHEELER

PROCESS DESCRIPTIONS

INPUTS-OUTPUTS SUMMARIES

FLOWSHEETS

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

REVIEW AND EVALUATION OF COAL LIQUEFACTION PROCESSES CONSIDERED FOR COMMERCIALIZATION AROUND 1995

CASE A: WESTINGHOUSE GASIFIER AND FISCHER-TROPSCH SYNTHESIS

Process Description

The flowsheet for this case is similar to Case IV (Winkler gasifier to F-T synthesis), with the change of Winkler to Westinghouse, and the deletion of the raw gas compression train. See attached flowsheet.

The Westinghouse oxygen blown gasifier is essentially a fluidized, ash agglomerating, single stage bed operation.

A 24 TPD pilot demonstration unit has been operated at Waltz Mill, Pennsylvania, with various types of coal. A Texas lignite has been reported successfully gasified at 1,000 kPa (130 psig) operating pressure.

Ground coal (1/4" x 0), dried to about 8% moisture, is charged to lockhoppers from where it is pneumatically conveyed into the gasifier via a central feed tube, located at the gasifier's bottom, where it is combusted and reacted with an oxidant stream of oxygen and steam.

The combustion of a portion of the incoming coal provides the heat necessary to devolatilze the remainder of the incoming coal to form a char and to react it with steam to form hydrogen and carbon monoxide.

As the bed of char circulates through the jet formed at the end of the central feed tube, the carbon in the char is consumed by the combustion and gasification, leaving ash rich particles.

The ash rich particles agglomerate into larger aggregates which defluidize from the bed and collect in the ash annulus discharging into lockhoppers, from where they are removed after being cooled by the uprising feed gas stream.

The raw gas leaving the gasifier passes through cyclones to recover the entrained particles which are recycled to the gasifier.

The raw gas, at about 1800°F, is passed through a heat recovery section followed by a wet particulate removal in a Venturi scrubber.

The remaining processing steps are indicated in the attached flowsheet and have been extensively described in other parts of this report.

The data developed for this case have been based on a proposed commercial plant, assumed operating at 4,240 kPa (600 psig) in the gasifiers. This is a large extrapolation of the 1,000 kPa attained in the PDU, and the risks associated with this scale-up have not been fully evaluated.

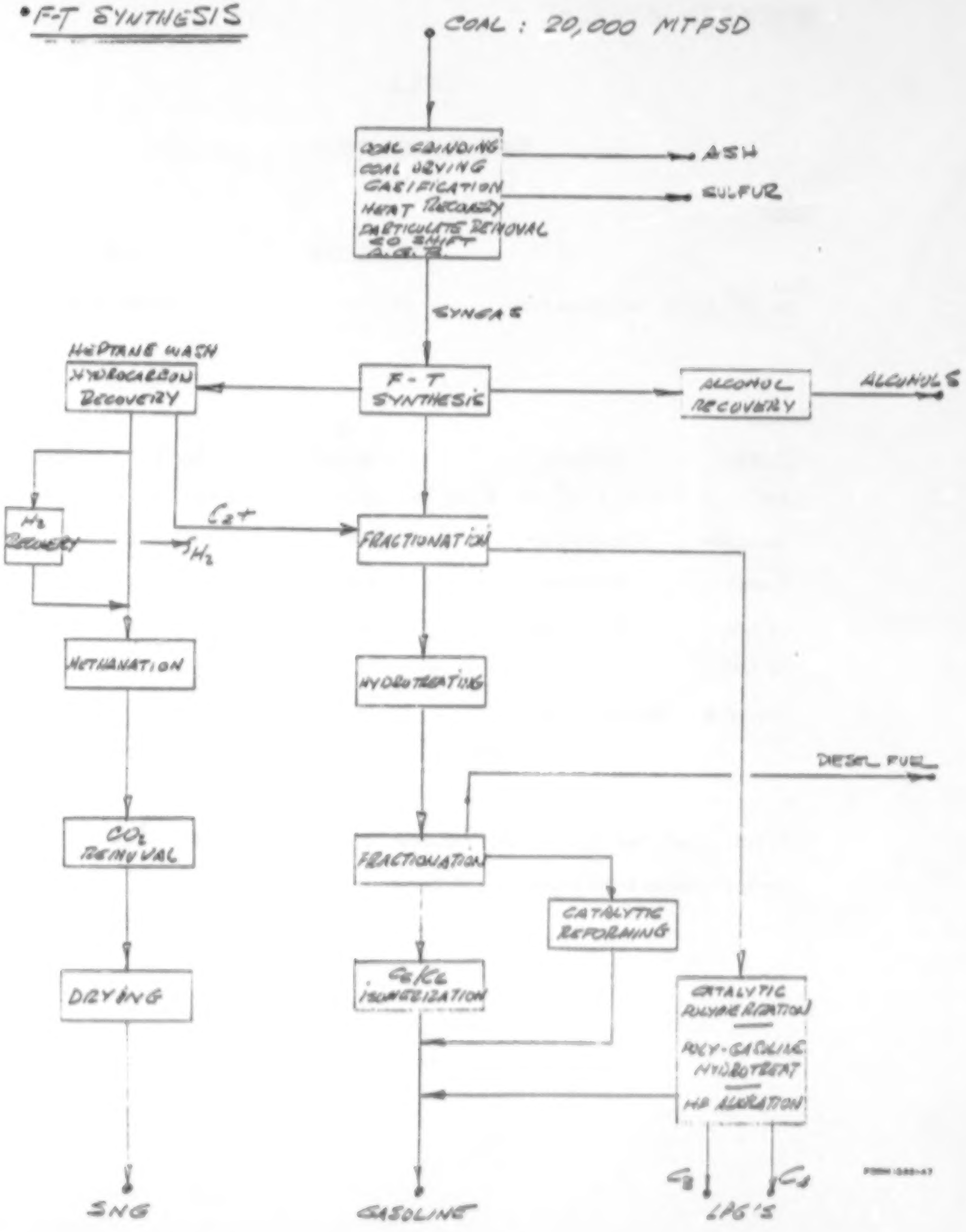
It should be noticed that due to relatively high pressure (4,240 kPa) and moderate temperature (1000°C) in the gasifiers, the expected generation of methane is substantial, about 10% by volume. This will require a reforming step if only liquid products are desired.

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N. J.

BY A.G. DATE 11/30/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. ______ OF ______

CHKD. BY ______ DATE ______ JOB NO. ______

- GASIFICATION
- F-T SYNTHESIS



FORM 1383-47

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE A

Input-Output Summary

Input

	HHV, kJ/kg	kJ/SD
Coal 20×10^6 Kg/SD (As Received)	15,774	3.1548×10^{11}

Output

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
SNG	2.8865×10^6 St. m^3/SD	36,843	1.0635×10^{11}	58.35
Gasoline	1,614 m^3/SD	3.386×10^7	5.4651×10^{10}	29.98
Diesel	308 m^3/SD	3.5188×10^7	1.0838×10^{10}	5.95
C_3 LPG	148 m^3/SD	2.5229×10^7	3.7339×10^9	2.05
C4 LPG	19 m^3/SD	2.7885×10^7	5.2981×10^8	0.29
Alcohols	194,320 kg/SD	3.1697×10^4	6.1593×10^9	3.38
		Totals	1.8226×10^{11}	100.00

F.O.E. (Fuel Oil Equivalent) @ 3.9836×10^7 Kj/m^3: 4,575 m^3/SD

Overall Thermal Efficiency: 57.8%

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE A-R

WESTINGHOUSE GASIFIER TO FISCHER-TROPSCH SYNTHESIS WITH SNG REFORMING

Process Description

The attached flowsheet indicates that the methanation train of the parent case (A) has been substituted for a SNG reforming train.

The purge stream from the F-T synthesis, which contains most of the methane, unreacted CO, H_2 and heavier hydrocarbons is washed with heptane for hydrocarbons recovery.

Most of the C_2's and heavier hydrocarbons in the purge stream are absorbed in the cold heptane (about-40°C) and returned to the fractionation train after being boiled off the heptane.

The methane containing stream leaving the heptane wash joins the light ends from the fractionation and the joined streams go through a hydrogen recovery step. Molecular sieves operating on the pressure swing mode effect the separation of the hydrogen from the other components.

The hydrogen barren stream leaving the hydrogen recovery plant is subjected to a partial oxidation in the SNG reforming plant where the hydrocarbons are converted into CO and H_2.

The net hydrogen left from the hydrogen recovery (after removing the the required amounts for the hydrotreating operations) is joined with the reformed stream and its CO_2 content adjusted (by removing the required excess) to meet the desired $H_2/(2\ CO + 3\ CO_2)$ ratio.

This adjusted stream constitutes the syngas recycle, which is fed back into the synthol reactors. The remaining steps of the flowsheet have been already described in the other parts of this report.

CASE A-R

Input-Output Summary

Input

	HHV, kJ/kg	kJ/SD
Coal 20×10^6 Kg/SD (As Received)	15,774	3.1548×10^{11}

Output

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
Gasoline	2,974 m^3/SD	3.386×10^7	1.007×10^{11}	74.27
Diesel Fuel	505 m^3/SD	3.5188×10^7	1.778×10^{10}	13.11
C_3 LPG	242 m^3/SD	2.5229×10^7	6.117×10^9	4.51
C_4 LPG	32 m^3/SD	2.7885×10^7	8.911×10^8	0.66
Alcohols	318,680 kg/SD	3.1697×10^4	1.010×10^{10}	7.45
		Totals	1.3559×10^{11}	100.00

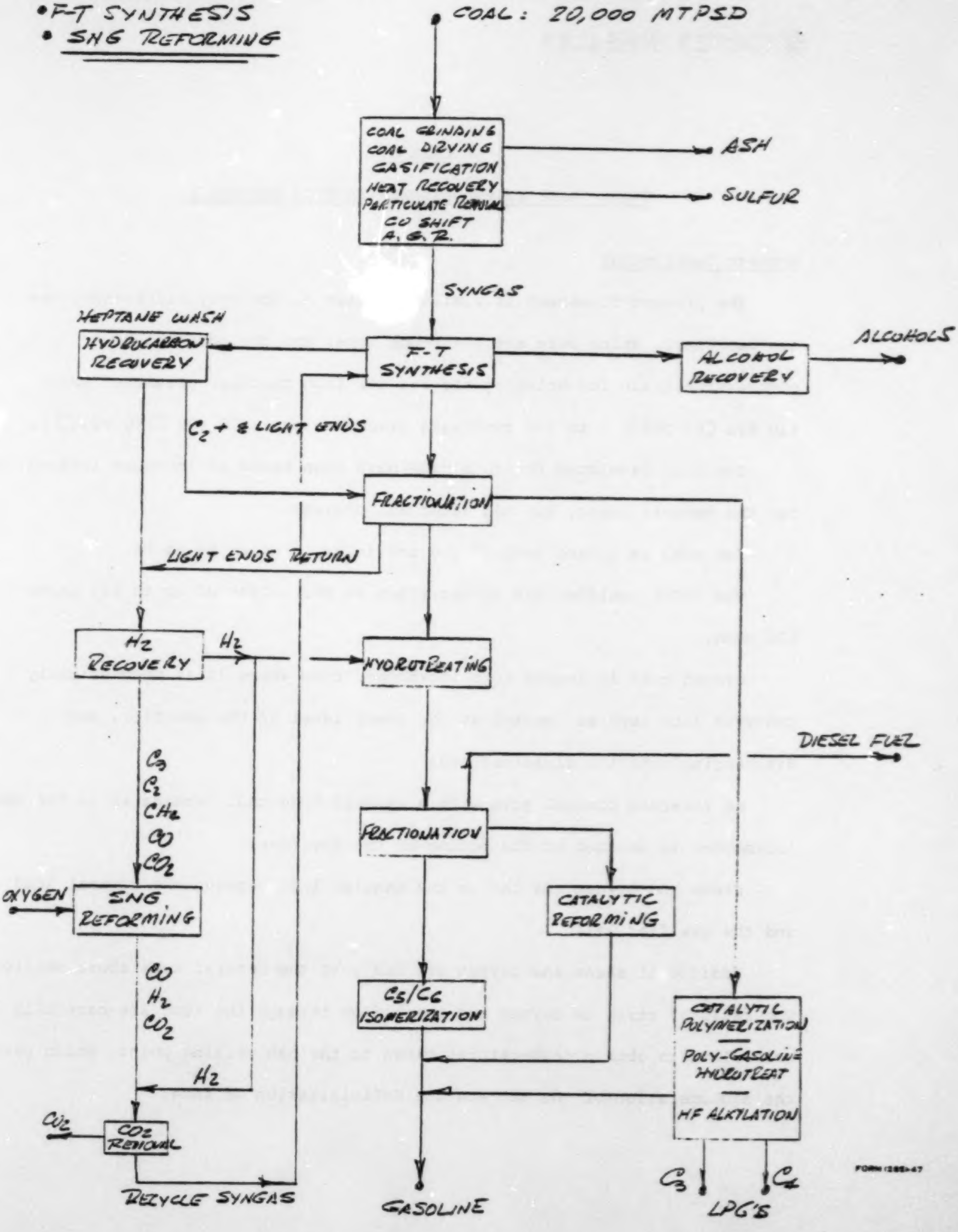
F.O.E. (Fuel Oil Equivalent) @ 3.9836 kJ/m^3: 3,404 m^3/SD

Overall Thermal Efficiency: 43%

BY A. G. DATE 11/30/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE GASIFICATION TO ALL LIQUID PRODUCTS JOB NO.

- GASIFICATION
- F-T SYNTHESIS
- SNG REFORMING



FORM 1285-47

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE B

U-GAS GASIFIER AND FISCHER-TROPSCH SYNTHESIS

Process Description

The process flowsheet is similar to Case A, the only differences are the gasifiers, which here are the U-Gas type, and the addition of a compression train for bringing the raw gas from gasifier pressure (about 620 kPa [90 psia]) to F-T synthesis pressure (ca. 2400 kPa [350 psia]).

The data developed for this case have been based on in-house information for the Memphis Light, Gas and Water Co. project.

The coal is ground to 1/4" x 0 and dried to about 5% to 8%.

The U-Gas gasifier can accept fines to the extent of up to 10% under 100 mesh.

Ground coal is loaded into lockhopper from where it is pneumatically conveyed into tuyeres located at the lower level of the gasifier, and discharging into the fluidized bed.

An inverted conical grid with a central tube that terminates in the ash lockhopper is located at the bottom of the gasifier.

Steam and oxygen are fed in the annulus left between the conical grid and the gasifier wall.

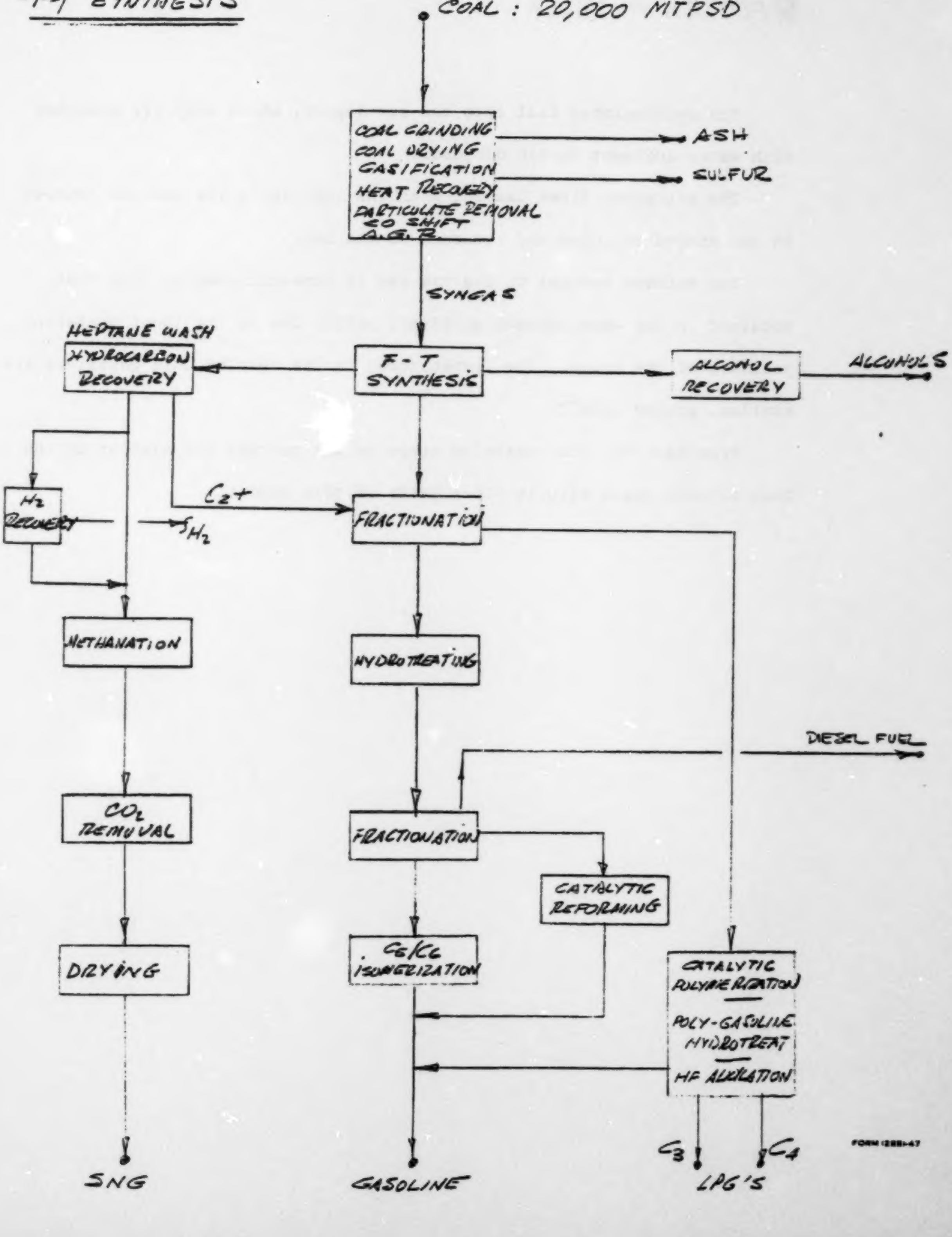
Additional steam and oxygen are fed into the central tube above mentioned. The ratio of steam to oxygen and their flow through the tube are carefully controlled to obtain temperatures close to the ash melting point, which permits the agglomeration of the ash and the defluidization of same.

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N. J.

BY A.G. DATE 11/30/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE JOB NO.

- GASIFICATION
- F-T SYNTHESIS



FORM 1285-47

The agglomerates fall into the ash hopper, where they are quenched with water and sent to ash disposal.

The entrained fines leaving with the gasifier's raw gas are removed in two staged cyclones and returned to the bed.

The methane content of the raw gas is somewhat smaller than that obtained in the Westinghouse gasifier, mainly due to the lower operating pressure of the U-Gas. The temperatures in the beds of both gasifiers are similar, around 1000°C.

From here on, the remaining steps of the process are similar to the ones already dealt with in other parts of this report.

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE B

INPUT-OUTPUT SUMMARY

Input

	HHV, kJ/kg	kJ/SD
Coal		
20 x 10^6 Kg/SD (As Received)	15,774	3.1548 x 10^{11}

Output

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
SNG	2.107 x 10^6 Std. m^3/SD	36,843	7.765 x 10^{10}	49.41
Gasoline	1,691 m^3/SD	3.386 x 10^7	5.726 x 10^{10}	36.42
Diesel Fuel	323 m^3/SD	3.5188 x 10^7	1.136 x 10^{10}	7.22
C_3 LPG	155 m^3/SD	2.5229 x 10^7	3.907 x 10^9	2.48
C_4 LPG	20 m^3/SD	2.7885 x 10^7	5.675 x 10^8	0.36
Alcohols	203,570 kg/SD	3.1697 x 10^4	6.453 x 10^9	4.11
		Totals	1.572 x 10^{11}	100.00

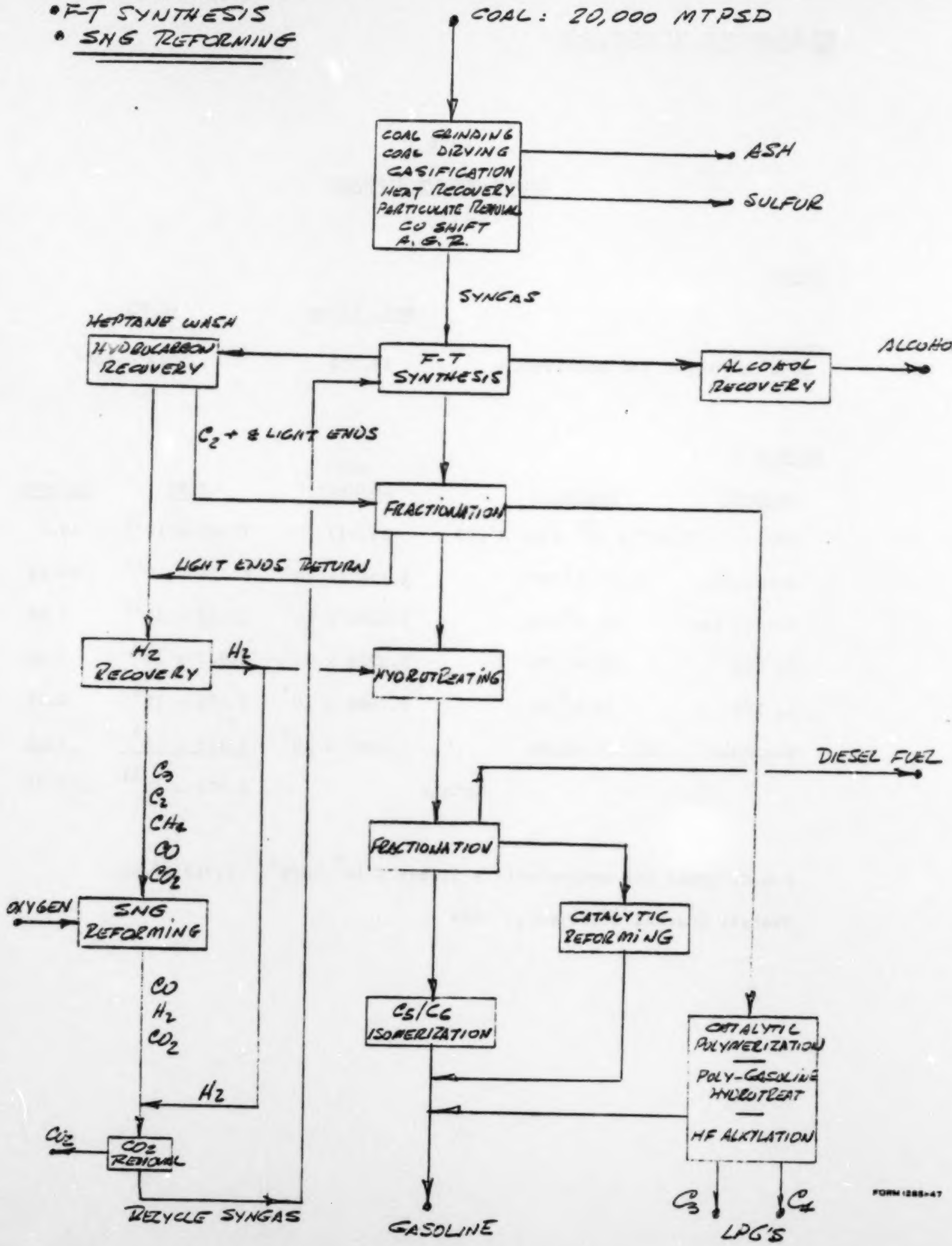
F.O.E. (Fuel Oil Equivalent) @ 3.9836 x 10^7 kJ/m^3: 3,946 m^3/SD

Overall Thermal Efficiency: 50%

BY A. G. DATE 11/30/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. ____ OF ____

CHKD. BY ____ DATE ____ GASIFICATION TO ALL LIQUID PRODUCTS JOB NO. ____

- GASIFICATION
- F-T SYNTHESIS
- SNG REFORMING



Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE B-R

U-GAS GASIFIER TO F-T WITH SNG REFORMING

Process Description

The attached flowsheet contains processing steps that have been described in other parts of this report and need no further explanation.

CASE B-R

INPUT-OUTPUT SUMMARY

Input

	HHV, kJ/kg	kJ/SD
Coal 20×10^6 kg/SD (As Received)	15,774	3.1548×10^{11}

Output

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
Gasoline	2,782 m^3/SD	3.386×10^7	9.418×10^{10}	74.27
Diesel Fuel	473 m^3/SD	3.5188×10^7	1.663×10^{10}	13.11
C_3 LPG	227 m^3/SD	2.5229×10^7	5.720×10^9	4.51
C_4 LPG	30 m^3/SD	2.7885×10^7	8.335×10^8	0.66
Alcohols	298,030 kg/SD	3.1697×10^4	9.447×10^9	7.45
		Totals	1.2681×10^{11}	100.00

F.O.E. (Fuel Oil Equivalent) @ 3.9836×10^7 kJ/m^3: 3,183 m^3/SD

Overall Thermal Efficiency: 40.2%

CASE C

SHELL-KOPPERS GASIFIER TO FISCHER-TROPSCH SYNTHESIS

Process Description

In the attached flowhsheet, the process steps have been already described in other parts of this report, and the only subject here that deserves consideration is the description of the gasification step.

The Shell-Koppers gasifier is similar in appearance to the Koppers-Totzek, the fundamental difference between them being the operating pressure. The S-K operates at about 2,860 kPa (400 psig).

It is claimed that practically any type of coal can be gasified, and the coals with ash content as high as 40% and high sulfur content are said to be gasified without any difficultly.

The flame temperature in the gasifier can be as high as 2000°C, and under these conditions the generation of tars is nil. Also, the methane content is relatively low, less than 1%, and CO_2 content is reported at about 5%. It produces a raw gas with a high $CO:H_2$ ratio, about 2, which would require a substantial shifting with its attendant steam requirements, for the F-T synthesis or methane synthesis, which require a $CO:H_2$ ratio of about 0.5.

The thermal efficiency is expected to be in the order ot about 75%.

As with the K-T, the high operating temperature slags the ash, which runs down to a quench tank.

A six ton per day pilot plant was in operation at the Shell Research Laboratory, in Amsterdam, since 1976.

A 165 ton per day built by Krupp-Koppers has been running since 1978, and it is claimed that the results have been successful enough to permit commercial scale-up.

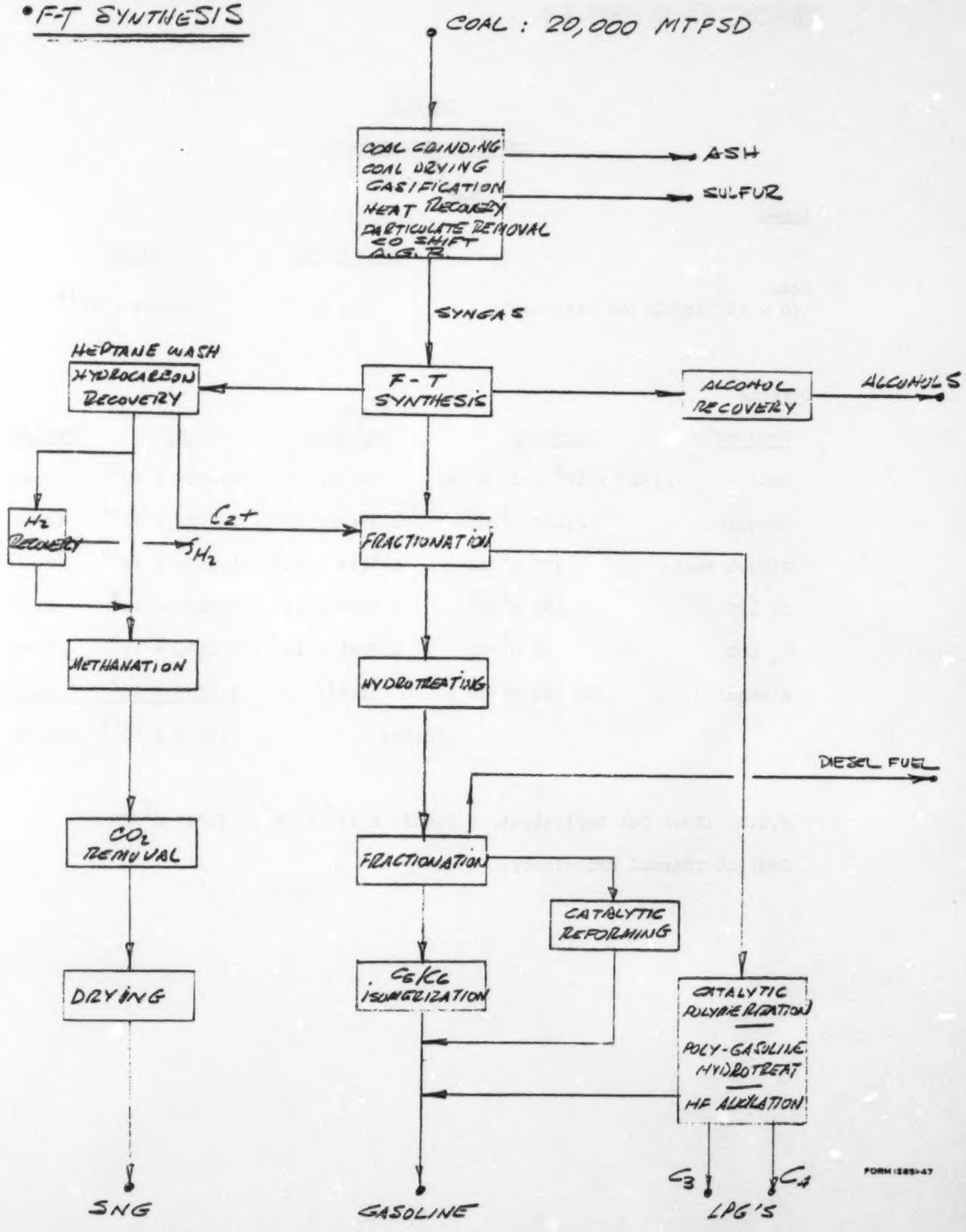
Two 1100 TPD prototype plants are in the design stage, and are planned to be operational in 1984-1985.

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N.J.

BY A.G. DATE 11/30/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE JOB NO.

- GASIFICATION
- F-T SYNTHESIS



FORM 1289-47

CASE C

INPUT-OUTPUT SUMMARY

Input

	HHV, kJ/kg	kJ/SD
Coal 20×10^6 kg/SD (As Received)	15,774	3.1548×10^{11}

Outputs

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
SNG	1.3827×10^6 Std. m^3/SD	36,843	4.8953×10^{10}	32.80
Gasoline	2,133 m^3/SD	3.386×10^7	7.2223×10^{10}	48.38
Diesel Fuel	407 m^3/SD	3.5188×10^7	1.4322×10^{10}	9.59
C_3 LPG	195 m^3/SD	2.5229×10^7	4.9197×10^9	3.29
C_4 LPG	26 m^3/SD	2.7885×10^7	7.2501×10^8	0.48
Alcohols	256,780 kg/SD	31,697	8.1392×10^9	5.46
		Totals	1.4928×10^{11}	100.00

F.O.E. (Fuel Oil Equivalent) @ 3.9836×10^7 kG/m^2: 3,748 m^3/SD

Overall Thermal Efficiency: 47.3%

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE C-R

SHELL-KOPPERS GASIFIERS TO F-T SYNTHESIS WITH SNG REFORMING

Process Description

Reference is made to the attached flowsheet. It should be remarked here that due to the very low methane content of the raw gas, the methane to be reformed is mainly generated in the F-T Synthol reactors.

When the product slate is desired to be maximized in gasoline content, it would be more convenient to process the syngas via the methanol to gasoline route, which is dealt with in the next case (C-1).

CASE C-R

INPUT-OUTPUT SUMMARY

Input

	HHV, kJ/kg	kJ/SD
Coal 20 x 10^6 kg/SD (As Received)	15,774	3.1548 x 10^{11}

Output

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
Gasoline	3,116 m^3/SD	3.386 x 10^7	1.055 x 10^{11}	74.25
Diesel Fuel	530 m^3/SD	3.5188 x 10^7	1.865 x 10^{10}	13.12
C_3 LPG	254 m^3/SD	2.5229 x 10^7	6.408 x 10^9	4.51
C_4 LPG	34 m^3/SD	2.7885 x 10^7	9.481 x 10^8	0.67
Alcohols	333,812 m^3/SD	31,697	1.058 x 10^{10}	7.45
		Totals	1.421 x 10^{11}	100.00

F.O.E. (Fuel Oil Equivalent) @ 3.9836 x 10^7 kJ/m^3: 3,567 m^3/SD

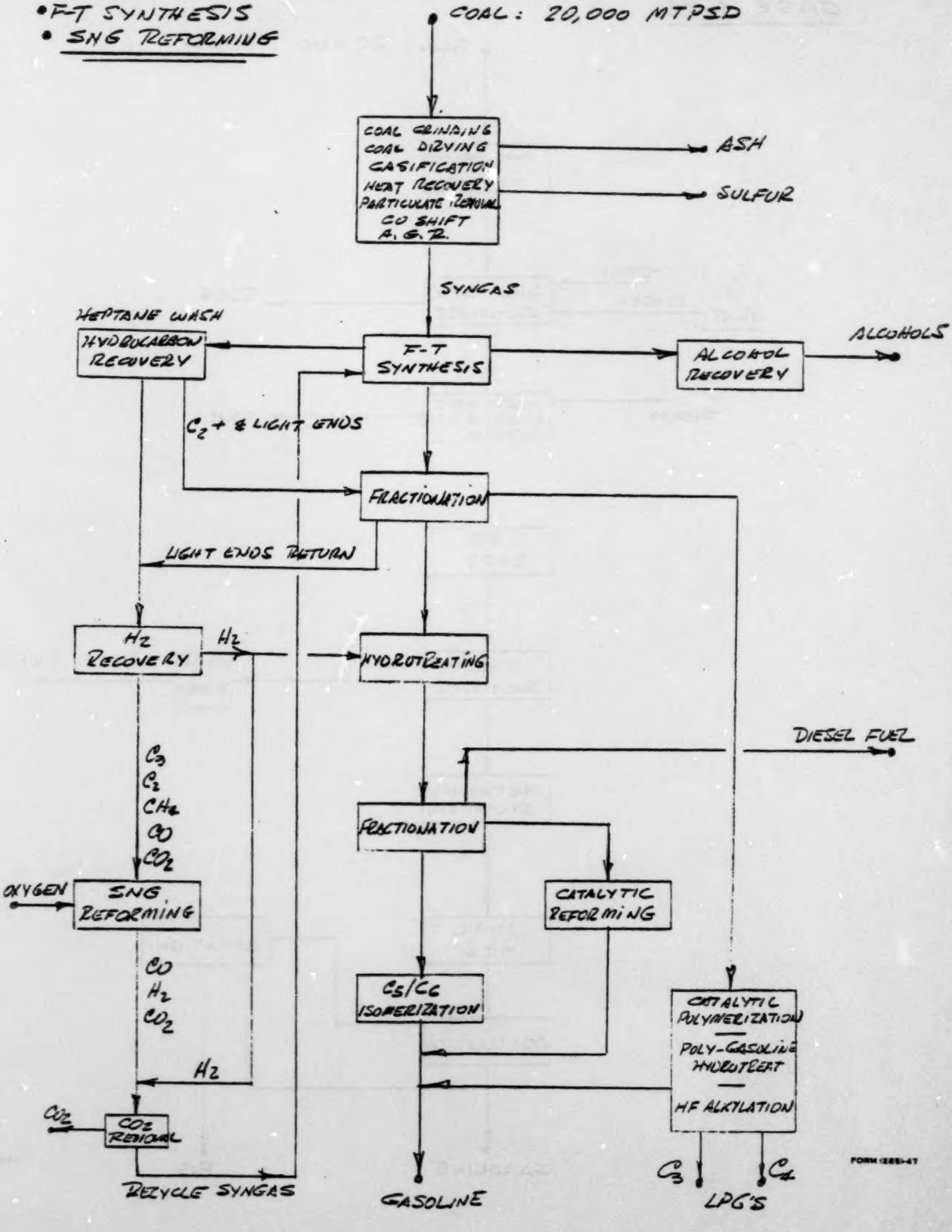
Overall Thermal Efficiency: 45%

FOSTER WHEELER ENERGY CORPORATION 110 SOUTH ORANGE AVE., LIVINGSTON, N. J.

BY A. G. DATE 11/30/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO. OF

CHKD. BY DATE GASIFICATION TO ALL LIQUID PRODUCTS JOB NO.

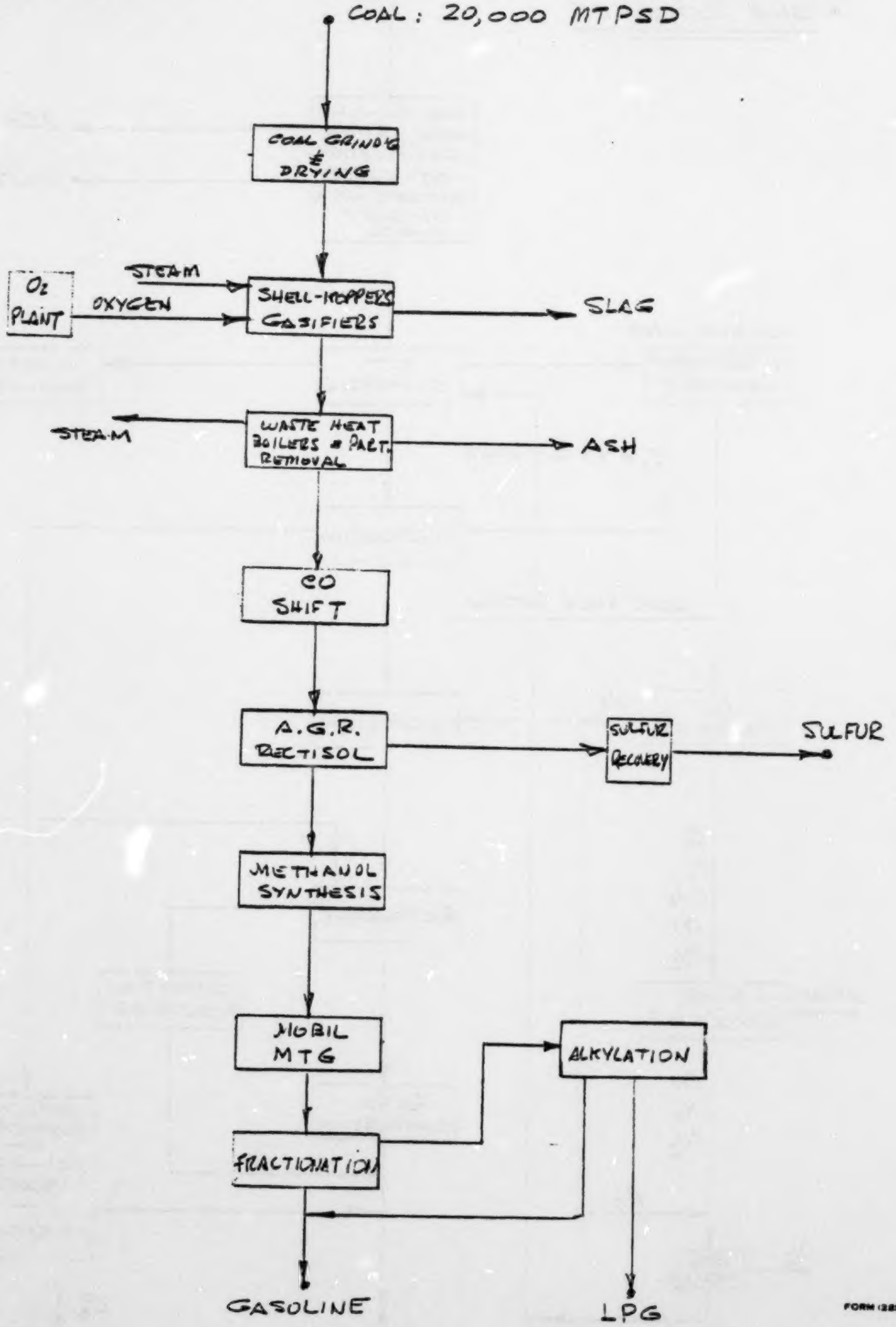
- GASIFICATION
- F-T SYNTHESIS
- SNG REFORMING



FORM 1285-47

BY G. G DATE 2/12/81 SUBJECT LIGNITE LIQUEFACTION SHEET NO ____ OF ____
CHKD. BY ____ DATE ____ JOB NO. B-23-5011

CASE C-1



CASE C-1

SHELL-KOPPERS GASIFIERS TO METHANOL TO GASOLINE (MTG)

Process Description

The Shell-Koppers gasifier does not generate any appreciable amounts of methane due to its high (about 1800° to 2000°C) operating temperature.

When the product slate is to be maximized in liquids, specially in gasoline content, a route requiring less processing steps is through methane synthesis followed by the Mobil methanol to gasoline process, as shown in the attached flowsheet.

The description of the process steps included here have been dealt with in other parts of this report.

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

CASE C-1

INPUT-OUTPUT SUMMARY

Input

	HHV, kJ/kg	kJ/SD
Coal 20 x 10^6 kG/SD (As Received)	15,774	3.1548 x 10^{11}

Output

Product	Quantity	HHV kJ/Unit	kJ/SD	Percent
Gasoline	3,629 m^3/SD	3.385 x 10^7	1.23 x 10^{11}	88.13
C_3 LPG	256 m^3/SD	2.5229 x 10^7	6.46 x 10^9	4.63
C_4 LPG	363 m^3/SD	2.7885 x 10^7	1.01 x 10^{10}	7.24
		Totals	1.40 x 10^{11}	100.00

F.O.E. (Fuel Oil Equivalent) @ 3.9836 x 10^7 kJ/m^3: 3,515 m^3/SD

Overall Thermal Efficiency: 44.4%

FW FOSTER WHEELER

COMPARATIVE INVESTMENT COSTS

COMPARATIVE INVESTMENT COSTS

The following investment costs were estimated for the various cases listed below. No claims for the accuracy of the absolute values is made, as only their relative values are of interest for the pruposes of this report.

Case A: Westinghouse Gasifier to Fischer-Tropsch Synthesis

$ 1,800 x 10^6

Case A-R: Westinghouse Gasifier to F-T with SNG Reforming

$ 2,050 x 10^6

The higher investment cost for this case is due to the increase in the oxygen plant capacity, the partial oxidation plant for the SNG reforming, the increase in the Boiler Plant, and the increase in the F-T train due to the addition of the recycle syngas from the SNG reforming.

Credit has been taken for the elimination of the methanation train.

Case B: U-Gas Gasifier to Fischer-Tropsch Synthesis

$ 1,650 x 10^6

Case B-R: U-Gas Gasifier to Fischer-Tropsch with SNG Reforming

$ 1,850 x 10^6

The higher investment cost for this case over the prior one is due to the reasons analogous to the A-R case.

Case C: Shell-Koppers Gasifier to F-T Synthesis

$\$ 1,700 \times 10^6$

Case C-R: Shell-Koppers Gasifier to F-T Synthesis with SNG Reforming

$\$ 1,850 \times 10^6$

Case C-1: Shell-Koppers Gasifier to Methanol to Gasoline

$\$ 1,600 \times 10^6$

An attempt was made to estimate the above costs under a consistent basis for all the cases considered.

COMMENTS

Job No.: B-28-8011
Date: December, 1981

FOSTER WHEELER

COMMENTS

The Westinghouse and U-Gas gasifiers are very similar in principle, both are single stage fluidized beds with the ash agglomerating feature.

There are, nevertheless, certain desimilarities, the most salient being:

1. The U-Gas consumes more steam than the Westinghouse. This may be due to the requirement of a higher fluidization velocity in the U-Gas case, steam is used for fluidization.
2. The Westinghouse requires the recycle of a sizable portion of the raw gas produced. The recycle is used as transport gas for coal feed and recycle of char fines.
3. The feeding of the coal into the gasifiers is effected in quite a different manner. The U-Gas feeds rapidly into the side of the bed, counting on the rapid heating and devolatilization by the fast mixing action of the coal jet with the circulating char.

The Westinghouse feeds axially through a vertical tube, resembling a spouted bed operation. The steam and oxygen are fed through the annulus between the coal feed tube and an externally concentric tube. It would thus appear that the steam-oxygen stream shrouds the reacting coal particles during their softening stage, thus improving the decoking process.

Both gasifiers generate sizable amounts of CH_4 due to the moderate temperature (about 1000°C). The pressure effect on methane generation would be more pronounced in Westinghouse, which produces a somewhat higher methane content due to its assumed higher operating pressure.

The U-Gas gasifier has been operated in the PDU (nominal 24 TPD) under atmospheric pressure, although some tests have been conducted at about 40 psig in order to gather data for the design of the proposed Memphis Industrial Fuel Gas Facility, which is planned to operate at 90 psia. It is not known if a lignite has been testes in the U-Gas. The Westinghouse PDU gasifier has been operated with oxygen at 130 psig, and from the data here obtained coupled with a cold flow model, a 600 psig plant design was extrapolated.

A Texas lignite was tested in the PDU, which would indicate that it may be suitable for the Estevan and Willow Bunch lignites.

There are no known plans at the present for the building of a large size plant for the Westinhouse gasifiers, although three demonstration-scale has been proposed.

One of the proposals involves a facility at Fairmount, West Virginia, to produce a medium BTU gas, with the gasifier operating at 220 psig, and feeding about 13 TPH of a Pittsburgh No. 2 seam. Another project at Keystone, Johnston, Pennsylvania, would be a coal to methanol plant, operating at 340 psig at the gasifiers and feeding about 14 TPH of Pittsburgh No. 2 steam coal.

The third one would be at Lake Chalres, Louisiana, to produce a billion BTU/day of medium-BTU gas for turbine fuel. The feed coal would be a Wyoming Sub-C, at the rate of about 100 TPH.

In general terms, it may be said that both gasifiers are still in the development stages and that problems associated with clinkering, deposition in cyclones and heat recovery surfaces, and with the fluidization regimes for the scaled-up versions of the PDU's must be resolved before reliable and continuous operation can be achieved.

As was pointed out above, both types of gasifiers produce a raw gas containing sizeable amounts of methane. It counter distinction, the Shell-Koppers gasifier, which is also in developmental stage, hardly generates any methane, mainly due to the high temperature of operation (1800°C to 2000°C).

All the three gasifiers - Westinghouse, U-Gas and Shell-Koppers - are intended for medium pressure operation (around 400 to 500 psig) which will allow the use of less compression power for the raw gas processing as well as diminish the size of the downstream equipment.

One of the problem areas with the S-K slagging gasifier could prove to be the refractory lining life. The pilot plant uses a quite complex water coaling, and it was reported that the placing of refractory lining was comtemplated.

The thermal efficiency of the S-K is reported to be in the order of 75%, it does not generate any tars or oils and would appear to be eminently suitable for the gasification of Saskatchewan lignites.

TABULATION OF SIGNIFICANT RESULTS

FOSTER WHEELER

Job No. B-28-8011
Date December, 1981

TABLE II

TABULATION OF SIGNIFICANT RESULTS PRESREENING STUDY

Case	A	A-R	B	B-R	C	C-R	C-1
Gasifier	Westinghouse	Westinghouse	U-Gas	U-Gas	Shell-Koppers	S-K	S-K
Liquefaction	K-T	F-T with SNG Reforming	F-T	F-T with SNG Reforming	F-T	F-T with SNG Reforming	Methanol to Gasoline
Gasifier Pressure, kPa	4,240	4,240	620	620	2,860	2,860	2,860
Overall Thermal Eff., %	57.8	43	50	40.2	47.3	45	44.4
Gasoline Yield, m^3/SD	1,614	2,974	1,691	2,782	2,133	3,116	3,629
Product Slate in Percent of Overall Efficiency							
SNG, %	58.35	-	49.41	-	32.80	-	-
Gasoline, %	29.98	74.27	36.42	74.27	48.38	74.25	88.13
Diesel Fuel, %	5.95	13.11	7.22	13.11	9.59	13.12	-
C_3 LPG, %	2.05	4.51	2.48	4.51	3.29	4.51	4.63
C_4 LPG, %	0.29	0.66	0.36	0.66	0.48	0.67	7.24
Alcohols, %	3.38	7.45	4.11	7.45	5.46	7.45	-
Fuel Oil Equivalent, m^3/SD	4,575	3,404	3,946	3,183	3,748	3,567	3,515
Syngas to Synthesis, kgm/hr	30,418	49,885	35,935	52,610	35,303	45,894	35,330
Investment Cost, MM$	1,800	2,050	1,650	1,850	1,700	1,850	1,600
Investment Cost per Daily m^3 of Gasoline, $	1,115,000	689,310	975,750	664,990	797,000	593,710	440,890
Investment Cost per Daily m^3 of F.O.E, $	393,440	602,230	418,140	581,211	453,580	518,640	455,190
Commercial Status, Year*	1,995	1,995	1,990	1,990	1,990	1,990	1,990

Note: Coal fed: 20×10^6 kg/SD (As Received)

* Estimated

REFERENCES

REFERENCES

1. L.A. Salvador et al, "The Westinghouse Coal Gasification Process," June, 1980.

2. J.D. Holmgren, C.E. Seglem, L.A. Salvador, M.W. Dyos, "The Westinghouse Gasification Program," EPRI Sponsored Conference, October 13-16, 1980.

3. C. Schwartz, K.K. Roth, M.D. Freir, "Application of the Westinghouse Coal Gasification Process," AIChE 1981 Summer National Meeting, Detroit, Michigan, August 16-19, 1981.

4. L.E. Bostwick and D. A. Hubbard, "Risk Analysis for Scale-Up of the Westinghouse Oxygen Blown Gasifier," D.O.E. AND G.R.I., March, 1981.

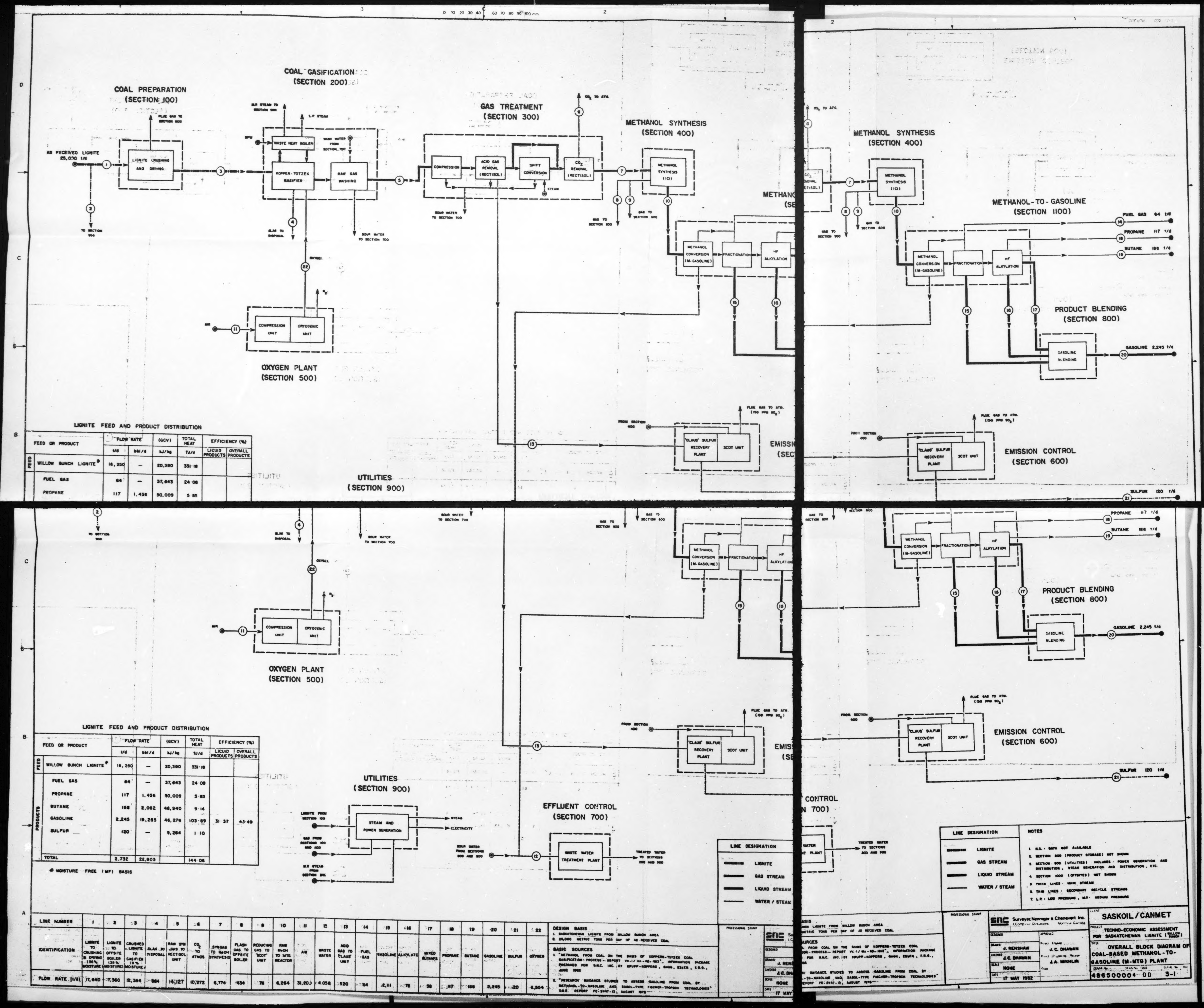
5. Max Schreiner, Mobil Research and Development Corp, D.O.E. FE-2447-13, August, 1981.

6. In-House Report on Memphis Industrial Fuel Gas Using U-Gas Gasification.

7. Chemical and Engineering News, August 1981, Gasification Units to Use Shell-Koppers Process.

8. J.T. Patel and J. W. Loeding, Institute of Gas Technology, U-Gas Demonstration Plant Program. Undated.

9. M. J. van der Burgt, Shell International Petroleum, The Hague, The Netherlands, Article in Hydrocarbon Processing, January, 1989, Clean Gas from Coal.


COAL PREPARATION (SECTION 100)
AS RECEIVED LIGNITE 25,000 t/d
LIGNITE CRUSHING AND DRYING
FLUE GAS TO SECTION 900
TO SECTION 900
COAL GASIFICATION (SECTION 200)
M.P. STEAM TO SECTION 900
L.P. STEAM
BFW
WASTE HEAT BOILER
KOPPER-TOTZEK GASIFIER
RAW GAS WASHING
WASH WATER FROM SECTION 700
SLAG TO DISPOSAL
SOUR WATER TO SECTION 700
OXYGEN
GAS TREATMENT (SECTION 300)
COMPRESSION
ACID GAS REMOVAL (RECTISOL)
SHIFT CONVERSION
CO2 REMOVAL (RECTISOL)
STEAM
CO2 TO ATM.
METHANOL SYNTHESIS (SECTION 400)
METHANOL SYNTHESIS (ICI)
GAS TO SECTION 900
GAS TO SECTION 600
METHANOL-TO-GASOLINE (SECTION 1100)
METHANOL CONVERSION (M-GASOLINE)
FRACTIONATION
HF ALKYLATION
FUEL GAS 64 t/d
PROPANE 117 t/d
BUTANE 186 t/d
PRODUCT BLENDING (SECTION 800)
GASOLINE BLENDING
GASOLINE 2,245 t/d
OXYGEN PLANT (SECTION 500)
AIR
COMPRESSION UNIT
CRYOGENIC UNIT
N2
EMISSION CONTROL (SECTION 600)
FROM SECTION 400
'CLAUS' SULFUR RECOVERY PLANT
SCOT UNIT
FLUE GAS TO ATM. (150 PPM SO2)
SULFUR 120 t/d
UTILITIES (SECTION 900)
LIGNITE FROM SECTION 100
GAS FROM SECTIONS 100 AND 1100
M.P. STEAM FROM SECTION 200
STEAM AND POWER GENERATION
STEAM
ELECTRICITY
EFFLUENT CONTROL (SECTION 700)
SOUR WATER FROM SECTIONS 200 AND 300
WASTE WATER TREATMENT PLANT
TREATED WATER TO SECTIONS 200 AND 900
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LIGNITE FEED AND PRODUCT DISTRIBUTION

	FEED OR PRODUCT	FLOW RATE		(GCV)	TOTAL HEAT	EFFICIENCY (%)	
		t/d	bbl/d	kJ/kg	TJ/d	LIQUID PRODUCTS	OVERALL PRODUCTS
FEED	WILLOW BUNCH LIGNITE *	16,250	—	20,380	331·18		
PRODUCTS	FUEL GAS	64	—	37,643	24·08	31·37	43·49
	PROPANE	117	1,456	50,009	5·85		
	BUTANE	188	2,062	48,940	9·14		
	GASOLINE	2,245	19,285	46,276	103·89		
	SULFUR	120	—	9,264	1·10		
	TOTAL	2,732	22,803		144·06		

* MOISTURE FREE (MF) BASIS

LINE NUMBER	1	2	3	4	5	6	7	8	9	10	11
IDENTIFICATION	LIGNITE TO CRUSHING & DRYING (35% MOISTURE)	LIGNITE TO OFFSITE BOILER (35% MOISTURE)	CRUSHED LIGNITE TO GASIFIER (8% MOISTURE)	SLAG TO DISPOSAL	RAW SYN GAS TO RECTISOL UNIT	CO_2 TO ATMOS.	SYNGAS TO MeOH SYNTHESIS	FLASH GAS TO OFFSITE BOILER	REDUCING GAS TO 'SCOT' UNIT	RAW MeOH TO MTG REACTOR	AIR
FLOW RATE (t/d)	17,640	7,360	12,384	984	14,127	10,272	6,774	434	78	6,264	31,200

LINE NUMBER	12	13	14	15	16	17	18	19	20	21	22
IDENTIFICATION	WASTE WATER	ACID GAS TO 'CLAUS' UNIT	FUEL GAS	GASOLINE	ALKYLATE	MIXED BUTANES	PROPANE	BUTANE	GASOLINE	SULFUR	OXYGEN
FLOW RATE (t/d)	4,058	520	64	2,111	76	58	117	186	2,245	120	6,504

DESIGN BASIS

1. SASKATCHEWAN LIGNITE FROM WILLOW BUNCH AREA
2. 25,000 METRIC TONS PER DAY OF AS RECEIVED COAL

BASIC SOURCES

1. "METHANOL FROM COAL ON THE BASIS OF KOPPERS-TOTZEK COAL GASIFICATION-PROCESS-REPORT VK-1/Mn-HD-1813", INFORMATION PACKAGE PREPARED FOR S.N.C. INC. BY KRUPP-KOPPERS, GmbH, ESSEN, F.R.G., JUNE 1982
2. "RESEARCH GUIDANCE STUDIES TO ASSESS GASOLINE FROM COAL BY METHANOL-TO-GASOLINE AND SASOL-TYPE FISCHER-TROPSCH TECHNOLOGIES" D.O.E. REPORT FE-2447-13, AUGUST 1978

LINE DESIGNATION	
▬	LIGNITE
▬	GAS STREAM
▬	LIQUID STREAM
—	WATER / STEAM

NOTES

1. N.A. : DATA NOT AVAILABLE
2. SECTION 800 (PRODUCT STORAGE) NOT SHOWN
3. SECTION 900 (UTILITIES) INCLUDES : POWER GENERATION AND DISTRIBUTION, STEAM GENERATION AND DISTRIBUTION, ETC.
4. SECTION 1000 (OFFSITES) NOT SHOWN
5. THICK LINES : MAIN STREAM
6. THIN LINES : SECONDARY RECYCLE STREAMS
7. L.P. : LOW PRESSURE, M.P. : MEDIUM PRESSURE

PROFESSIONAL STAMP

SNC Surveyer, Nenniger & Chenevert Inc.
1 Complexe Desjardins Montreal Canada

DESIGNED
DRAWN J. RENSHAW
CHECKED J.C. DHAWAN
SCALE NONE
DATE 17 MAY 1982

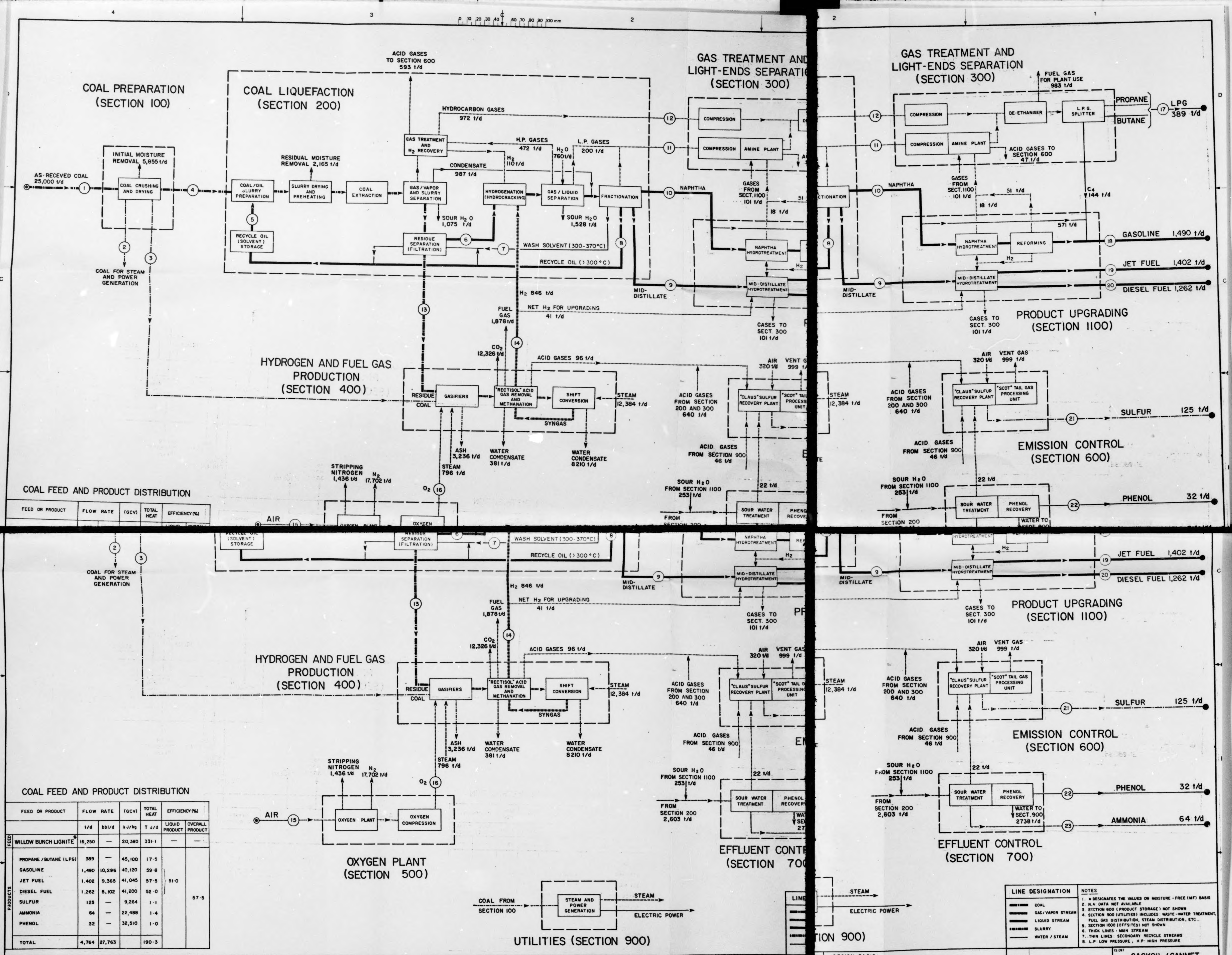
APPROVED
Project Engineer J.C. DHAWAN
Project Engineering Manager J.A. MIKHLIN
Client

CLIENT SASKOIL / CANMET

PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)

TITLE OVERALL BLOCK DIAGRAM OF COAL-BASED METHANOL-TO-GASOLINE (M-MTG) PLANT

CONTR No. 486500004 DRAWING CODE 00 SERIAL No. 3-1 Rev


COAL PREPARATION (SECTION 100)
AS-RECEVED COAL 25,000 t/d
INITIAL MOISTURE REMOVAL 5,855 t/d
COAL CRUSHING AND DRYING
COAL FOR STEAM AND POWER GENERATION
COAL LIQUEFACTION (SECTION 200)
COAL/OIL SLURRY PREPARATION
RECYCLE OIL (SOLVENT) STORAGE
RESIDUAL MOISTURE REMOVAL 2,165 t/d
SLURRY DRYING AND PREHEATING
COAL EXTRACTION
GAS/VAPOR AND SLURRY SEPARATION
ACID GASES TO SECTION 600 593 t/d
GAS TREATMENT AND H2 RECOVERY
HYDROCARBON GASES 972 t/d
H.P. GASES 472 t/d
CONDENSATE 987 t/d
H2 110 t/d
HYDROGENATION (HYDROCRACKING)
H2O 760 t/d
L.P. GASES 200 t/d
GAS/LIQUID SEPARATION
FRACTIONATION
SOUR H2O 1,075 t/d
SOUR H2O 1,528 t/d
RESIDUE SEPARATION (FILTRATION)
WASH SOLVENT (300-370°C)
RECYCLE OIL (> 300°C)
NAPHTHA
MID-DISTILLATE
H2 846 t/d
NET H2 FOR UPGRADING 41 t/d
FUEL GAS 1,878 t/d
CO2 12,326 t/d
ACID GASES 96 t/d
HYDROGEN AND FUEL GAS PRODUCTION (SECTION 400)
RESIDUE
COAL
GASIFIERS
"RECTISOL" ACID GAS REMOVAL AND METHANATION
SHIFT CONVERSION
SYNGAS
STEAM 12,384 t/d
ASH 3,236 t/d
STEAM 796 t/d
WATER CONDENSATE 381 t/d
WATER CONDENSATE 8210 t/d
STRIPPING NITROGEN 1,436 t/d
N2 17,702 t/d
O2
AIR
OXYGEN PLANT
OXYGEN COMPRESSION
OXYGEN PLANT (SECTION 500)
GAS TREATMENT AND LIGHT-ENDS SEPARATION (SECTION 300)
COMPRESSION
DE-ETHANISER
L.P.G. SPLITTER
AMINE PLANT
FUEL GAS FOR PLANT USE 983 t/d
PROPANE
BUTANE
LPG 389 t/d
ACID GASES TO SECTION 600 47 t/d
C4 144 t/d
GASES FROM SECT. 1100 101 t/d
51 t/d
18 t/d
57 t/d
NAPHTHA HYDROTREATMENT
REFORMING
H2
MID-DISTILLATE HYDROTREATMENT
GASOLINE 1,490 t/d
JET FUEL 1,402 t/d
DIESEL FUEL 1,262 t/d
GASES TO SECT. 300 101 t/d
PRODUCT UPGRADING (SECTION 1100)
AIR 320 t/d
VENT GAS 999 t/d
ACID GASES FROM SECTION 200 AND 300 640 t/d
"CLAUS" SULFUR RECOVERY PLANT
"SCOT" TAIL GAS PROCESSING UNIT
SULFUR 125 t/d
ACID GASES FROM SECTION 900 46 t/d
EMISSION CONTROL (SECTION 600)
SOUR H2O FROM SECTION 1100 253 t/d
22 t/d
FROM SECTION 200 2,603 t/d
SOUR WATER TREATMENT
PHENOL RECOVERY
PHENOL 32 t/d
WATER TO SECT. 900 2738 t/d
AMMONIA 64 t/d
EFFLUENT CONTROL (SECTION 700)
COAL FROM SECTION 100
STEAM AND POWER GENERATION
STEAM
ELECTRIC POWER
UTILITIES (SECTION 900)
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COAL FEED AND PRODUCT DISTRIBUTION

	FEED OR PRODUCT	FLOW RATE		(GCV)	TOTAL HEAT	EFFICIENCY (%)	
		t/d	bbl/d	kJ/kg	T J/d	LIQUID PRODUCT	OVERALL PRODUCT
FEED	WILLOW BUNCH LIGNITE*	16,250	—	20,380	331·1	—	—
PRODUCTS	PROPANE / BUTANE (LPG)	389	—	45,100	17·5		57·5
	GASOLINE	1,490	10,296	40,120	59·8	51·0	
	JET FUEL	1,402	9,365	41,045	57·5		
	DIESEL FUEL	1,262	8,102	41,200	52·0		
	SULFUR	125	—	9,264	1·1		
	AMMONIA	64	—	22,488	1·4		
	PHENOL	32	—	32,510	1·0		
	TOTAL	4,764	27,763		190·3		

LINE NO	IDENTIFICATION	FLOW RATE (t/d)
1	AS-RECEIVED COAL	25,000
2	COAL FOR STEAM AND POWER	1,162
3	COAL TO GASIFIERS	1,333
4	COAL TO EXTRACTOR	16,650
5	SOLVENT (> 300°C)	33,827
6	COAL EXTRACT AND SOLVENT	37,531
7	WASH SOLVENT (300-370°C)	1,284
8	RECYCLE SOLVENT (> 300°C)	32,360
9	MID-DISTILLATE (200-300°C)	2,780
10	NAPHTHA (C_5-200°C)	1,390
11	LOW-PRESSURE SOUR GASES	420
12	HYDRO-CARBON GASES	972
13	RESIDUAL CHAR	7,333
14	TOTAL MAKE-UP H_2	887
15	AIR	24,307
16	OXYGEN (98 mol % O_2)	5,169
17	PROPANE BUTANE (LPG)	389
18	GASOLINE	1,490
19	JET FUEL	1,402
20	DIESEL FUEL	1,262
21	SULFUR	125
22	PHENOL	32
23	AMMONIA	64
24		
25		

DESIGN BASIS

1. SASKATCHEWAN LIGNITE FROM WILLOW BUNCH AREA.
2. 25,000 METRIC TONS PER DAY OF AS-RECEIVED COAL.
3. COAL CONVERSION IN EXTRACTOR: 65% MAF BASIS

BASIC SOURCES

1. NCB REPORT, "COAL LIQUEFACTION FOR TRANSPORT FUEL USING LIQUID SOLVENT EXTRACTION TECNOLOGY OF THE NATIONAL COAL BOARD", FEB 1981
2. NCB COMMUNICATIONS TO SNC, "THE PROCESSING OF SASKATCHEWAN LIGNITES BY THE NCB'S LIQUID SOLVENT EXTRACTION PROCESS", NCB REF No DR 3 000, DR 496/2, OCT. 1981, AND DR 496/2, 7/3, JAN. 25, 1982

LINE DESIGNATION	
▬ ▪ ▬	COAL
▬▬▬	GAS / VAPOR STREAM
▬▬	LIQUID STREAM
▪▪▪	SLURRY
—	WATER / STEAM

NOTES

1. * DESIGNATES THE VALUES ON MOISTURE-FREE (MF) BASIS
2. N.A: DATA NOT AVAILABLE
3. SECTION 800 (PRODUCT STORAGE) NOT SHOWN
4. SECTION 900 (UTILITIES) INCLUDES: WASTE-WATER TREATMENT, FUEL GAS DISTRIBUTION, STEAM DISTRIBUTION, ETC...
5. SECTION 1000 (OFFSITES) NOT SHOWN
6. THICK LINES: MAIN STREAM
7. THIN LINES: SECONDARY RECYCLE STREAMS
8. L.P: LOW PRESSURE; H.P: HIGH PRESSURE

SNC Surveyer, Nenniger & Chênevert Inc.
1, Complexe Desjardins, Montreal, Canada

DESIGNED		APPROVED
DRAWN	M. AIOUTZ	Project Engineer F.A. ASHRAF
CHECKED	F.A. ASHRAF	Project Engineering Manager J.A. MIKHLIN
SCALE	NONE	D: 1
DATE	27 JAN. 1982	

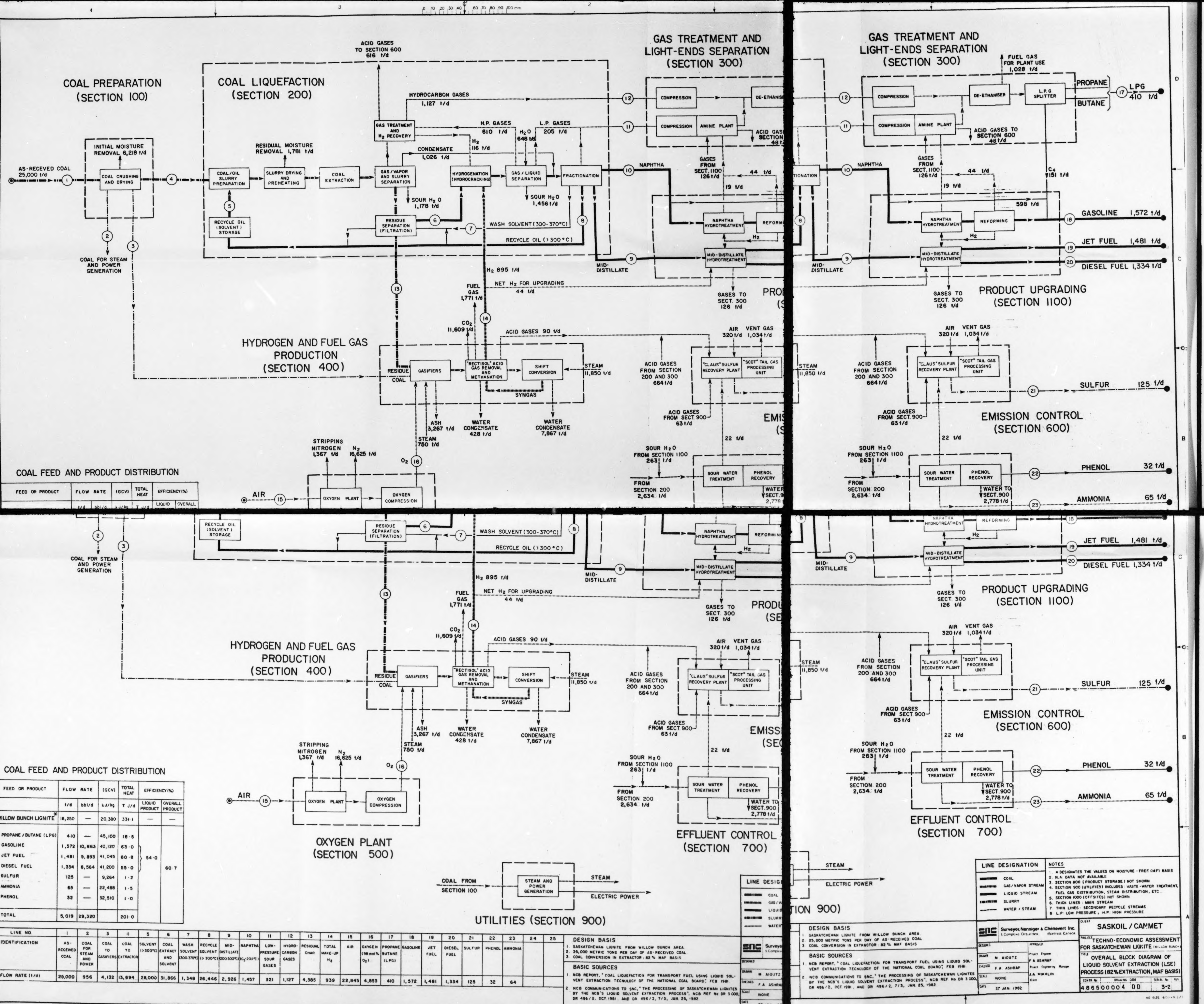
CLIENT SASKOIL / CANMET

PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)

TITLE OVERALL BLOCK DIAGRAM OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS (65% EXTRACTION, MAF BASIS)

CONTR No	DRAWING CODE			SERIAL No	REV
48650000	4	00		3-1	

A0 SIZE 4000-16 E/1


COAL PREPARATION (SECTION 100)
AS-RECEVED COAL 25,000 t/d
INITIAL MOISTURE REMOVAL 6,218 t/d
COAL CRUSHING AND DRYING
COAL FOR STEAM AND POWER GENERATION
COAL LIQUEFACTION (SECTION 200)
ACID GASES TO SECTION 600 616 t/d
COAL/OIL SLURRY PREPARATION
RESIDUAL MOISTURE REMOVAL 1,781 t/d
SLURRY DRYING AND PREHEATING
COAL EXTRACTION
GAS/VAPOR AND SLURRY SEPARATION
GAS TREATMENT AND H2 RECOVERY
HYDROCARBON GASES 1,127 t/d
H.P. GASES 610 t/d
L.P. GASES 205 t/d
H2O 648 t/d
CONDENSATE 1,026 t/d
H2 116 t/d
HYDROGENATION (HYDROCRACKING)
GAS / LIQUID SEPARATION
FRACTIONATION
SOUR H2O 1,178 t/d
SOUR H2O 1,456 t/d
RECYCLE OIL (SOLVENT) STORAGE
RESIDUE SEPARATION (FILTRATION)
WASH SOLVENT (300-370°C)
RECYCLE OIL (> 300°C)
NAPHTHA
MID-DISTILLATE
H2 895 t/d
NET H2 FOR UPGRADING 44 t/d
FUEL GAS 1,771 t/d
CO2 11,609 t/d
ACID GASES 90 t/d
HYDROGEN AND FUEL GAS PRODUCTION (SECTION 400)
RESIDUE COAL
GASIFIERS
"RECTISOL" ACID GAS REMOVAL AND METHANATION
SHIFT CONVERSION
SYNGAS
STEAM 11,850 t/d
ASH 3,267 t/d
STEAM 750 t/d
WATER CONDENSATE 428 t/d
WATER CONDENSATE 7,867 t/d
STRIPPING NITROGEN 1,367 t/d
N2 16,625 t/d
O2
AIR
OXYGEN PLANT
OXYGEN COMPRESSION
OXYGEN PLANT (SECTION 500)
GAS TREATMENT AND LIGHT-ENDS SEPARATION (SECTION 300)
COMPRESSION
DE-ETHANISER
L.P.G. SPLITTER
FUEL GAS FOR PLANT USE 1,028 t/d
PROPANE
BUTANE
LPG 410 t/d
AMINE PLANT
ACID GASES TO SECTION 600 48 t/d
GASES FROM SECT. 1100 126 t/d
44 t/d
19 t/d
C4 151 t/d
598 t/d
NAPHTHA HYDROTREATMENT
REFORMING
H2
MID-DISTILLATE HYDROTREATMENT
GASOLINE 1,572 t/d
JET FUEL 1,481 t/d
DIESEL FUEL 1,334 t/d
GASES TO SECT. 300 126 t/d
PRODUCT UPGRADING (SECTION 1100)
AIR 320 t/d
VENT GAS 1,034 t/d
ACID GASES FROM SECTION 200 AND 300 664 t/d
"CLAUS" SULFUR RECOVERY PLANT
"SCOT" TAIL GAS PROCESSING UNIT
ACID GASES FROM SECT. 900 63 t/d
22 t/d
SULFUR 125 t/d
EMISSION CONTROL (SECTION 600)
SOUR H2O FROM SECTION 1100 263 t/d
FROM SECTION 200 2,634 t/d
SOUR WATER TREATMENT
PHENOL RECOVERY
WATER TO SECT. 900 2,778 t/d
PHENOL 32 t/d
AMMONIA 65 t/d
EFFLUENT CONTROL (SECTION 700)
COAL FROM SECTION 100
STEAM AND POWER GENERATION
STEAM
ELECTRIC POWER
UTILITIES (SECTION 900)
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COAL FEED AND PRODUCT DISTRIBUTION

FEED OR PRODUCT	FLOW RATE		(GCV)	TOTAL HEAT	EFFICIENCY (%)	
	t/d	bbl/d	kJ/kg	T J/d	LIQUID PRODUCT	OVERALL PRODUCT
WILLOW BUNCH LIGNITE*	16,250	—	20,380	331·1	—	—
PROPANE / BUTANE (LPG)	410	—	45,100	18·5		60·7
GASOLINE	1,572	10,863	40,120	63·0	54·0	
JET FUEL	1,481	9,893	41,045	60·8		
DIESEL FUEL	1,334	8,564	41,200	55·0		
SULFUR	125	—	9,264	1·2		
AMMONIA	65	—	22,488	1·5		
PHENOL	32	—	32,510	1·0		
TOTAL	5,019	29,320		201·0		

LINE NO	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
IDENTIFICATION	AS-RECEIVED COAL	COAL FOR STEAM AND POWER	COAL TO GASIFIERS	COAL TO EXTRACTOR	SOLVENT (>300°C)	COAL EXTRACT AND SOLVENT	WASH SOLVENT (300-370°C)	RECYCLE SOLVENT (> 300°C)	MID-DISTILLATE (200-300°C)	NAPHTHA (C_5-200°C)	LOW-PRESSURE SOUR GASES	HYDRO-CARBON GASES	RESIDUAL CHAR	TOTAL MAKE-UP H_2	AIR	OXYGEN (98 mol % O_2)	PROPANE BUTANE (LPG)	GASOLINE	JET FUEL	DIESEL FUEL	SULFUR	PHENOL	AMMONIA		
FLOW RATE (t/d)	25,000	956	4,132	13,694	28,000	31,866	1,348	26,446	2,926	1,457	321	1,127	4,385	939	22,845	4,853	410	1,572	1,481	1,334	125	32	64		

LINE DESIGNATION

▬ ▪ ▬	COAL
▪▪▪▪▪	GAS / VAPOR STREAM
▬▬	LIQUID STREAM
▪▬▪▬	SLURRY
– – –	WATER / STEAM

NOTES
1. * DESIGNATES THE VALUES ON MOISTURE - FREE (MF) BASIS
2. N.A: DATA NOT AVAILABLE
3. SECTION 800 (PRODUCT STORAGE) NOT SHOWN
4. SECTION 900 (UTILITIES) INCLUDES : WASTE - WATER TREATMENT, FUEL GAS DISTRIBUTION, STEAM DISTRIBUTION, ETC...
5. SECTION 1000 (OFFSITES) NOT SHOWN
6. THICK LINES : MAIN STREAM
7. THIN LINES : SECONDARY RECYCLE STREAMS
8. L.P: LOW PRESSURE ; H.P: HIGH PRESSURE

DESIGN BASIS
1. SASKATCHEWAN LIGNITE FROM WILLOW BUNCH AREA
2. 25,000 METRIC TONS PER DAY OF AS-RECEIVED COAL
3. COAL CONVERSION IN EXTRACTOR : 82 % MAF BASIS

BASIC SOURCES
1. NCB REPORT, "COAL LIQUEFACTION FOR TRANSPORT FUEL USING LIQUID SOLVENT EXTRACTION TECNOLOGY OF THE NATIONAL COAL BOARD", FEB 1981
2. NCB COMMUNICATIONS TO SNC, "THE PROCESSING OF SASKATCHEWAN LIGNITES BY THE NCB'S LIQUID SOLVENT EXTRACTION PROCESS", NCB REF No DR 3 000, DR 496 / 2, OCT 1981, AND DR 496 / 2, 7/3, JAN. 25, 1982

SNC Surveyer, Nenniger & Chênevert Inc.
1, Complexe Desjardins, Montreal, Canada

DESIGNED
DRAWN M. AIOUTZ
CHECKED F. A. ASHRAF
SCALE NONE
DATE 27 JAN 1982

APPROVED
Project Engineer F. A. ASHRAF
Project Engineering Manager J. A. MIKHLIN
Client

CLIENT SASKOIL / CANMET

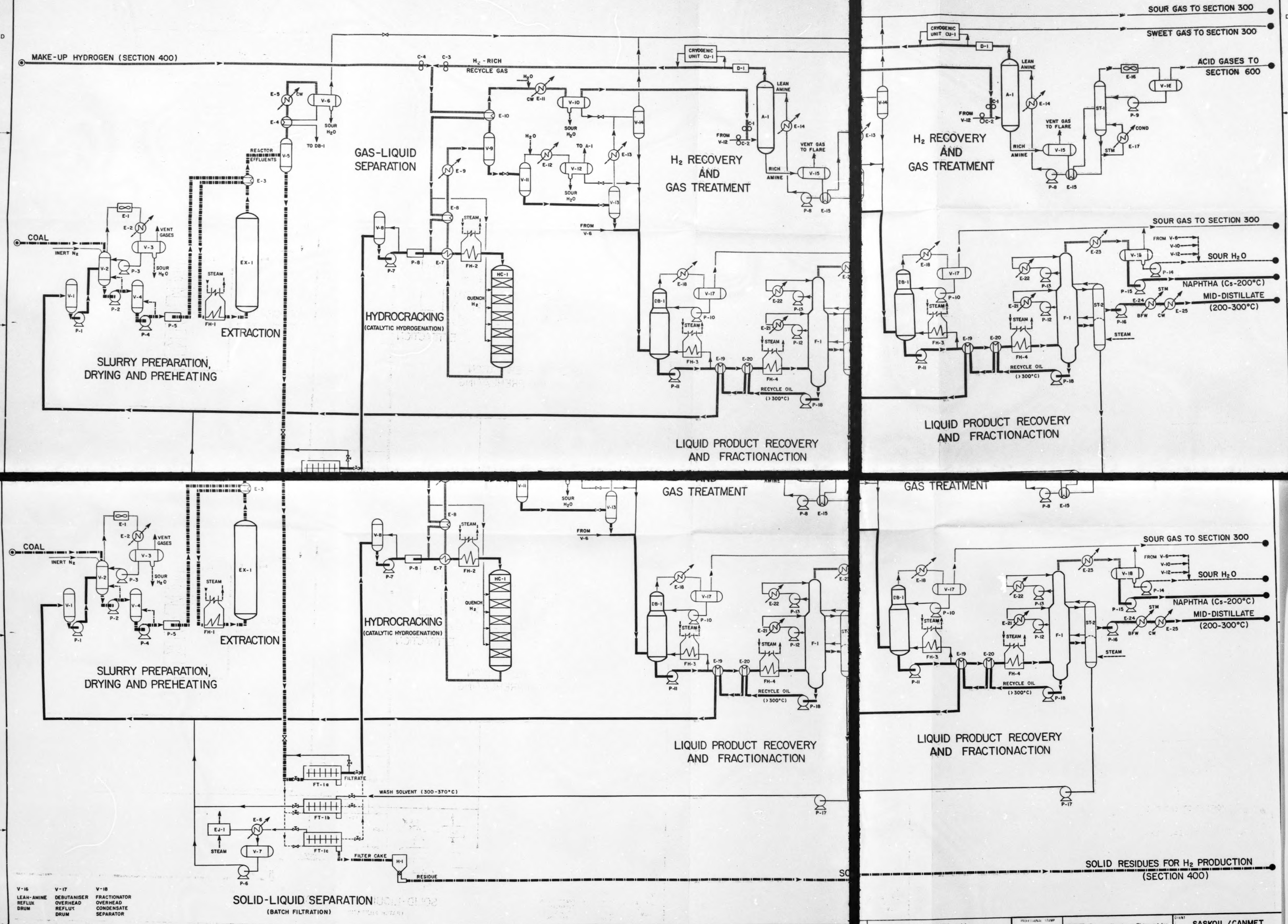
PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)

TITLE OVERALL BLOCK DIAGRAM OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS (82% EXTRACTION, MAF BASIS)

CONTR No 4865 DRAWING CODE 00004 DD SERIAL No 3-2 REV

A0 SIZE

A-1	HIGH-PRESSURE AMINE ABSORBER
CU-1	CRYOGENIC UNIT
D-1	H_2-RICH GAS DRIER (MOLECULAR SIEVES)
DB-1	DEBUTANISER
EJ-1	STEAM-EJECTOR SYSTEMS
EX-1	EXTRACTION REACTOR
F-1	ATMOSPHERIC DISTILLATION TOWER
FT-1	DISC FILTERS
H-1	RESIDUE HOPPER
HD-1	CATALYTIC HYDROCRACKER
ST-1	AMINE STRIPPER
ST-2	FRACTIONATOR SIDE-STREAM STEAM STRIPER
V-1	SOLVENT DRUM
V-2	SLURRY PREPARATION TANK
V-3	SLURRY PREPARATION OVERHEAD DRUM
V-4	SLURRY HOLDING TANK
V-5	REACTOR-EFFLUENT HOT SEPARATOR
V-6	REACTOR EFFLUENT OVERHEAD DRUM
V-7	RECOVERED SOLVENT DRUM
V-8	CONDENSATE (EXTRACT) DRUM
V-9	HIGH-PRESSURE HIGH-TEMP SEPARATOR
V-10	HIGH-PRESSURE LOW-TEMP CONDENSATE SEPARATOR
V-11	INT.-PRESSURE HIGH-TEMP FLASH DRUM
V-12	INT.-PRESSURE LOW-TEMP CONDENSATE SEPARATOR
V-13	LOW-PRESSURE HIGH-TEMP FLASH DRUM
V-14	LOW-PRESSURE LOW-TEMP CONDENSATE SEPARATOR
V-15	RICH-AMINE FLAS-DRUM
V-16	LEAN-AMINE REFLUX DRUM
V-17	DEBUTANISER OVERHEAD REFLUX DRUM
V-18	FRACTIONATOR OVERHEAD CONDENSATE SEPARATOR



LINE DESIGNATION

▬ ▪ ▬	COAL / SOLID RESIDUE
▪▪▪	SLURRY
┅┅┅	GAS / VAPOR STREAM
▬▬▬	LIQUID STREAM
- - - -	WATER / STEAM

EQUIPMENT DESIGNATION

- A - ABSORBER
- C - COMPRESSOR
- CU - CRYOGENIC UNIT
- D - DRIER
- DB - DEBUTANISER
- E - HEAT EXCHANGER
- EJ - EJECTOR
- E - EXTRACTION REACTOR
- FH - FIRED HEATER
- FT - FILTER
- H - HOPPER
- HC - HYDROCRACKER
- P - PUMP
- ST - STEAM STRIPPER
- V - VESSEL (S)

DESIGN BASIS

1. WILLOW BUNCH LIGNITE FROM SASKATCHEWAN (CANADA)
2. 25,000 METRIC TONS PER DAY OF AS-RECEIVED COAL TO THE PLANT.

BASIC SOURCES

1. NCB REPORT, "COAL LIQUEFACTION FOR TRANSPORTATION FUEL USING THE LIQUID SOLVENT EXTRACTION TECHNOLOGY OF THE U.K. NATIONAL COAL BOARD, FEB 1981
2. NCB COMMUNICATIONS TO SNC, "THE PROCESSING OF SASKATCHEWAN LIGNITES BY THE NCB'S LIQUID SOLVENT EXTRACTION PROCESS", NCB REF. N. DR 3000, DR 496/2, OCT. 1981; AND DR 496/2, 7/3, JAN. 25, 1982

FLOWRATES, t/d

EXTRACTION EFFICIENCY (MAF)	65 %	82 %
COAL	16,650	13,694
MAKE-UP HYDROGEN	887	939
PROCESS H_2O	760	648
SOUR GAS	420	321
SWEET GAS	972	1,127
ACID GASES	593	616
SOUR H_2O	2,603	2,634
NAPHTHA (C_5-200°C)	1,390	1,457
MID-DISTILLATE (200-300°C)	2,780	2,926
RESIDUES	7,333	4,385
VENT GASES	2,165	1,781

FLOWRATES, t/d

EXTRACTION EFFICIENCY (MAF)	65 %	82 %
COAL	16,650	13,694
MAKE-UP HYDROGEN	887	939
PROCESS H_2O	760	648
SOUR GAS	420	321
SWEET GAS	972	1,127
ACID GASES	593	616
SOUR H_2O	2,603	2,634
NAPHTHA (C_5-200°C)	1,390	1,457
MID-DISTILLATE (200-300°C)	2,780	2,926
RESIDUES	7,333	4,385
VENT GASES	2,165	1,781

PROFESSIONAL STAMP

SNC Surveyer, Nenniger & Chenevert Inc.
1 Complexe Desjardins Montreal Canada

DESIGNED
DRAWN M. AIOUTZ, J. LACHAINE
CHECKED
SCALE NONE
DATE JAN. 27, 1982

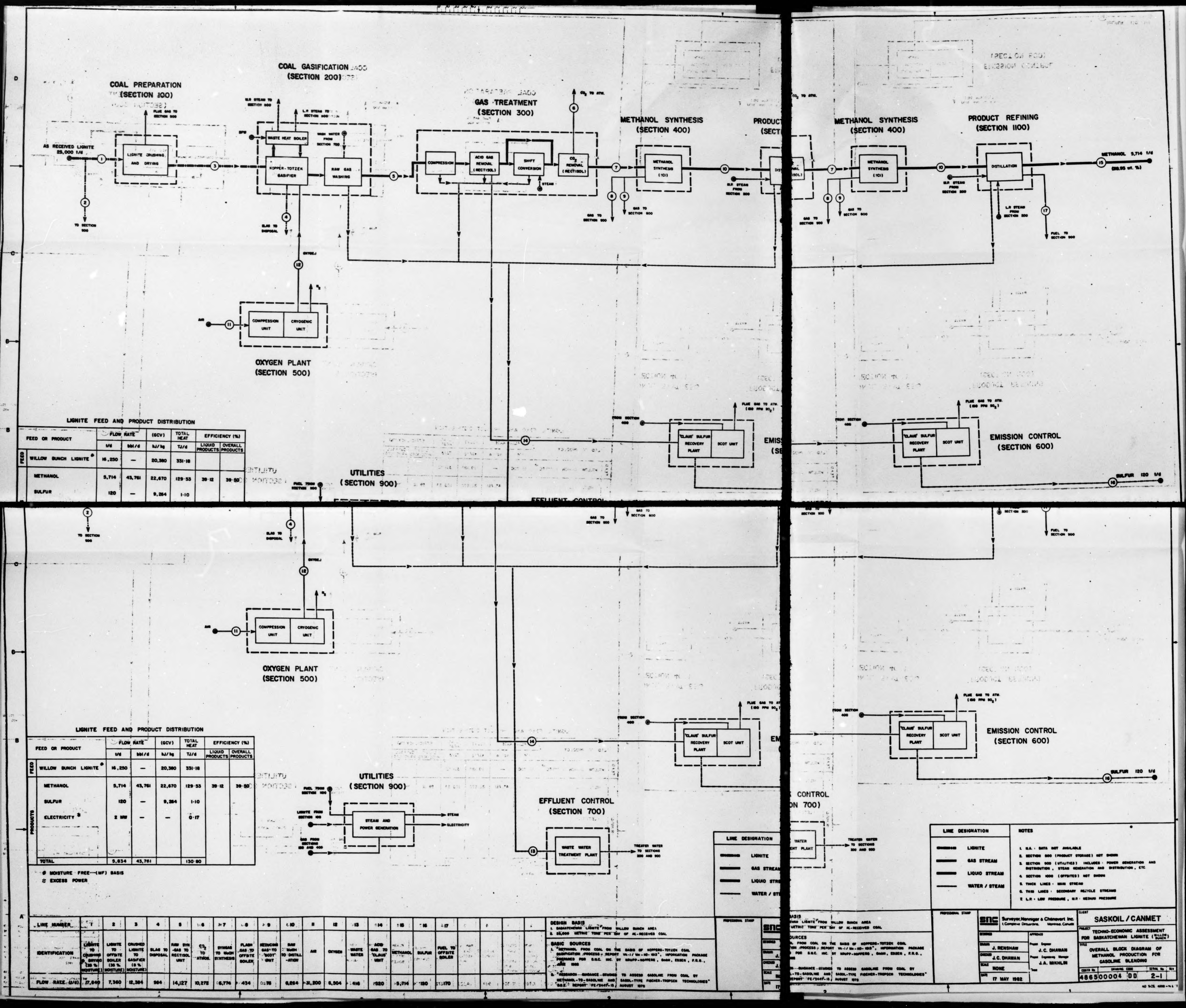
APPROVED
Project Engineer - F. A. ASHRAF
Project Engineering Manager J. A. MIKHLIN
Client

CLIENT SASKOIL / CANMET

PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)

TITLE PROCESS FLOW DIAGRAM OF COAL LIQUEFACTION (SECTION 200) OF LIQUID SOLVENT EXTRACTION (LSE) PROCESS

CONTR No	DRAWING CODE	SERIAL No	REV
4865	4 DD	3-3	


COAL PREPARATION (SECTION 100)
AS RECEIVED LIGNITE 25,000 t/d
LIGNITE CRUSHING AND DRYING
FLUE GAS TO SECTION 900
TO SECTION 900
COAL GASIFICATION (SECTION 200)
M.P. STEAM TO SECTION 900
L.P. STEAM TO SECTION 900
BFW
WASTE HEAT BOILER
KOPPERS-TOTZEK GASIFIER
WASH WATER FROM SECTION 700
RAW GAS WASHING
SLAG TO DISPOSAL
OXYGEN
GAS TREATMENT (SECTION 300)
COMPRESSION
ACID GAS REMOVAL (RECTISOL)
SHIFT CONVERSION
STEAM
CO_2 REMOVAL (RECTISOL)
CO_2 TO ATM.
GAS TO SECTION 900
METHANOL SYNTHESIS (SECTION 400)
METHANOL SYNTHESIS (ICI)
M.P. STEAM FROM SECTION 900
PRODUCT REFINING (SECTION 1100)
DISTILLATION
L.P. STEAM FROM SECTION 900
FUEL TO SECTION 900
METHANOL 5,714 t/d (99.95 wt. %)
AIR
COMPRESSION UNIT
CRYOGENIC UNIT
N_2
OXYGEN PLANT (SECTION 500)
FROM SECTION 400
'CLAUS' SULFUR RECOVERY PLANT
SCOT UNIT
FLUE GAS TO ATM. (150 PPM SO_2)
EMISSION CONTROL (SECTION 600)
SULFUR 120 t/d
UTILITIES (SECTION 900)
FUEL FROM SECTION 800
LIGNITE FROM SECTION 100
GAS FROM SECTIONS 100 AND 400
STEAM AND POWER GENERATION
STEAM
ELECTRICITY
EFFLUENT CONTROL (SECTION 700)
WASTE WATER TREATMENT PLANT
TREATED WATER TO SECTIONS 200 AND 900
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LIGNITE FEED AND PRODUCT DISTRIBUTION

FEED OR PRODUCT		FLOW RATE		(GCV)	TOTAL HEAT	EFFICIENCY (%)	
		t/d	bbl/d	kJ/kg	TJ/d	LIQUID PRODUCTS	OVERALL PRODUCTS
FEED	WILLOW BUNCH LIGNITE ⌀	16,250	—	20,380	331·18		
PRODUCTS	METHANOL	5,714	43,761	22,670	129·53	39·12	39·50
	SULFUR	120	—	9,284	1·10		
	ELECTRICITY ✲	2 MW	—	—	0·17		
	TOTAL	5,834	43,761		130·80		

⌀ MOISTURE FREE—(MF) BASIS
✲ EXCESS POWER

LINE NUMBER	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
IDENTIFICATION	LIGNITE TO CRUSHING & DRYING (36 % MOISTURE)	LIGNITE TO OFFSITE BOILER (36 % MOISTURE)	CRUSHED LIGNITE TO GASIFIER (8 % MOISTURE)	SLAG TO DISPOSAL	RAW SYN-GAS TO RECTISOL UNIT	CO_2 TO ATMOS.	SYNGAS TO MeOH SYNTHESIS	FLASH GAS TO OFFSITE BOILER	REDUCING GAS TO 'SCOT' UNIT	RAW MeOH TO DISTILLATION	AIR	OXYGEN	WASTE WATER	ACID GAS TO 'CLAUS' UNIT	METHANOL	SULFUR	FUEL TO OFFSITE BOILER
FLOW RATE (t/d)	17,640	7,360	12,384	984	14,127	10,272	6,774	434	76	6,264	31,200	8,504	416	520	5,714	120	170

DESIGN BASIS

1. SASKATCHEWAN LIGNITE FROM WILLOW BUNCH AREA
2. 25,000 METRIC TONS PER DAY OF AS-RECEIVED COAL

BASIC SOURCES

1. "METHANOL FROM COAL ON THE BASIS OF KOPPERS-TOTZEK COAL GASIFICATION PROCESS - REPORT VA-1/Wo-KO-1613", INFORMATION PACKAGE PREPARED FOR S.N.C. INC. BY KRUPP-KOPPERS, GmbH, ESSEN, F.R.G., JUNE 1980
2. "RESEARCH - GUIDANCE - STUDIES TO ASSESS GASOLINE FROM COAL BY METHANOL-TO-GASOLINE AND SASOL-TYPE FISCHER-TROPSCH TECHNOLOGIES" D.O.E. REPORT FE-2447-13, AUGUST 1978

LINE DESIGNATION	
▬	LIGNITE
▬	GAS STREAM
▬	LIQUID STREAM
—	WATER / STEAM

NOTES

1. N.A. : DATA NOT AVAILABLE
2. SECTION 800 (PRODUCT STORAGE) NOT SHOWN
3. SECTION 900 (UTILITIES) INCLUDES : POWER GENERATION AND DISTRIBUTION, STEAM GENERATION AND DISTRIBUTION, ETC.
4. SECTION 1000 (OFFSITES) NOT SHOWN
5. THICK LINES : MAIN STREAM
6. THIN LINES : SECONDARY RECYCLE STREAMS
7. L.P. : LOW PRESSURE, M.P. : MEDIUM PRESSURE

PROFESSIONAL STAMP

SNC Surveyer, Nenniger & Chênevert Inc.
1, Complexe Desjardins, Montreal, Canada

DESIGNED	
DRAWN	J. RENSHAW
CHECKED	J.C. DHAWAN
SCALE	NONE
DATE	17 MAY 1982
APPROVED	
Project Engineer	J.C. DHAWAN
Project Engineering Manager	J.A. MIKHLIN
Client	

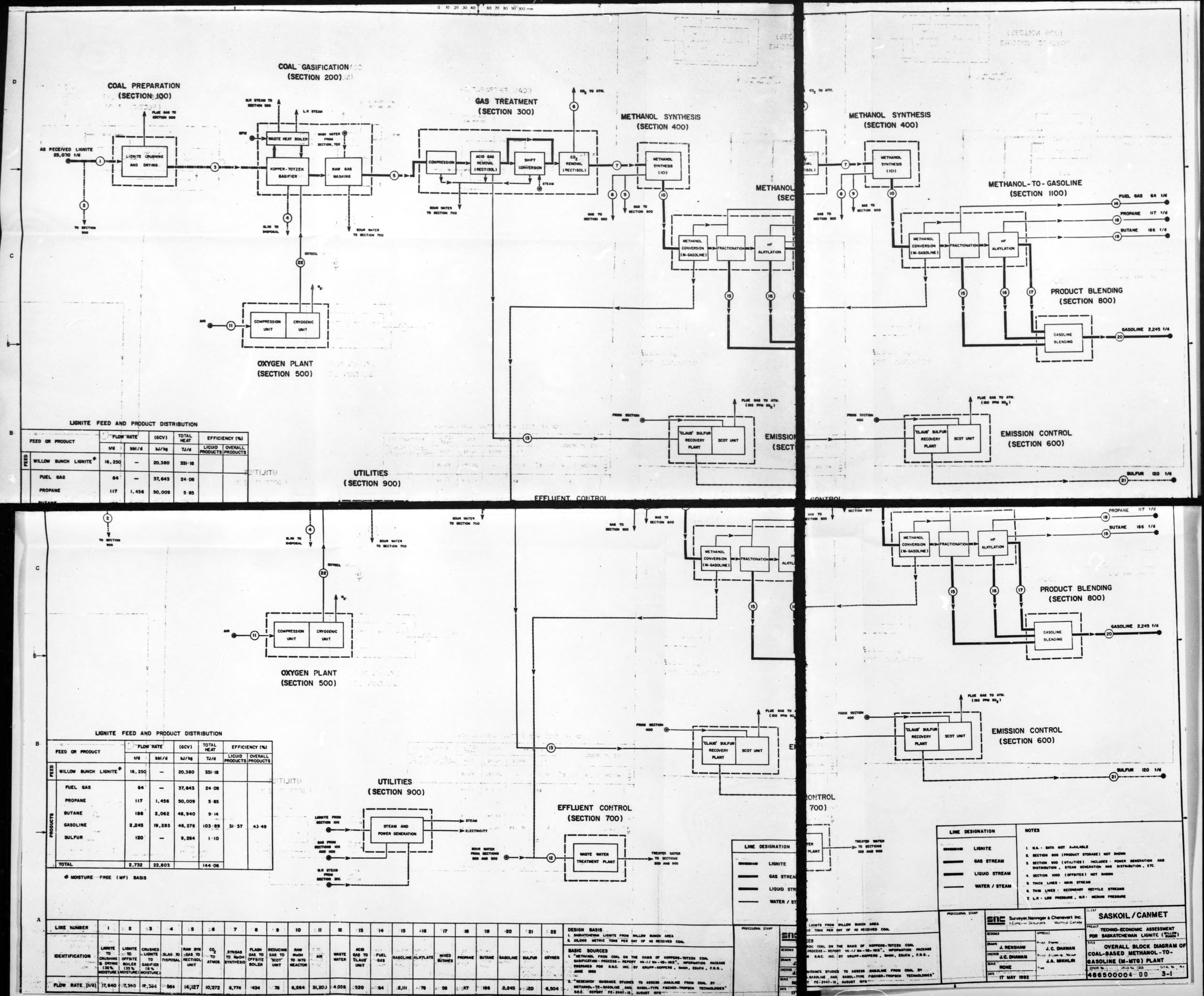
CLIENT SASKOIL / CANMET

PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)

TITLE OVERALL BLOCK DIAGRAM OF METHANOL PRODUCTION FOR GASOLINE BLENDING

CONTR No	DRAWING CODE		SERIAL No	REV
4865	00004	DD	2-1	

A0 SIZE 4000-1%L



LIGNITE FEED AND PRODUCT DISTRIBUTION

	FEED OR PRODUCT	FLOW RATE		(GCV)	TOTAL HEAT	EFFICIENCY (%)	
		t/d	bbl/d	kJ/kg	TJ/d	LIQUID PRODUCTS	OVERALL PRODUCTS
FEED	WILLOW BUNCH LIGNITE ⊕	16,250	—	20,380	331·18		
PRODUCTS	FUEL GAS	64	—	37,643	24·08	31·37	43·49
	PROPANE	117	1,456	50,009	5·85		
	BUTANE	186	2,062	48,940	9·14		
	GASOLINE	2,245	19,285	46,276	103·89		
	SULFUR	120	—	9,264	1·10		
	TOTAL	2,732	22,803		144·06		

⊕ MOISTURE FREE (MF) BASIS

LINE NUMBER	IDENTIFICATION	FLOW RATE (t/d)
1	LIGNITE TO CRUSHING & DRYING (36% MOISTURE)	17,640
2	LIGNITE TO OFFSITE BOILER (35% MOISTURE)	7,360
3	CRUSHED LIGNITE TO GASIFIER (8% MOISTURE)	12,384
4	SLAG TO DISPOSAL	864
5	RAW SYN-GAS TO RECTISOL UNIT	14,127
6	CO_2 TO ATMOS.	10,272
7	SYNGAS TO MeOH SYNTHESIS	6,774
8	FLASH GAS TO OFFSITE BOILER	434
9	REDUCING GAS TO 'SCOT' UNIT	78
10	RAW MeOH TO MTG REACTOR	6,264
11	AIR	31,200
12	WASTE WATER	4,058
13	ACID GAS TO 'CLAUS' UNIT	520
14	FUEL GAS	64
15	GASOLINE	2,111
16	ALKYLATE	76
17	MIXED BUTANES	58
18	PROPANE	117
19	BUTANE	186
20	GASOLINE	2,245
21	SULFUR	120
22	OXYGEN	6,504

DESIGN BASIS

1. SASKATCHEWAN LIGNITE FROM WILLOW BUNCH AREA
2. 25,000 METRIC TONS PER DAY OF AS RECEIVED COAL

BASIC SOURCES

1. "METHANOL FROM COAL ON THE BASIS OF KOPPERS-TOTZEK COAL GASIFICATION PROCESS – REPORT VK-1/Mb-KD-1013", INFORMATION PACKAGE PREPARED FOR S.N.C. INC. BY KRUPP-KOPPERS, GmbH, ESSEN, F.R.G., JUNE 1982
2. "RESEARCH GUIDANCE STUDIES TO ASSESS GASOLINE FROM COAL BY METHANOL-TO-GASOLINE AND SASOL-TYPE FISCHER-TROPSCH TECHNOLOGIES" D.O.E. REPORT FE-2447-13, AUGUST 1978

LINE DESIGNATION	
▬▬	LIGNITE
━━	GAS STREAM
━━	LIQUID STREAM
──	WATER / STEAM

NOTES

1. N.A. : DATA NOT AVAILABLE
2. SECTION 800 (PRODUCT STORAGE) NOT SHOWN
3. SECTION 900 (UTILITIES) INCLUDES : POWER GENERATION AND DISTRIBUTION, STEAM GENERATION AND DISTRIBUTION, ETC.
4. SECTION 1000 (OFFSITES) NOT SHOWN
5. THICK LINES : MAIN STREAM
6. THIN LINES : SECONDARY RECYCLE STREAMS
7. L.P. : LOW PRESSURE, M.P. : MEDIUM PRESSURE

PROFESSIONAL STAMP

SNC Surveyer, Nenniger & Chenevert Inc.
1 Complexe Desjardins Montreal Canada

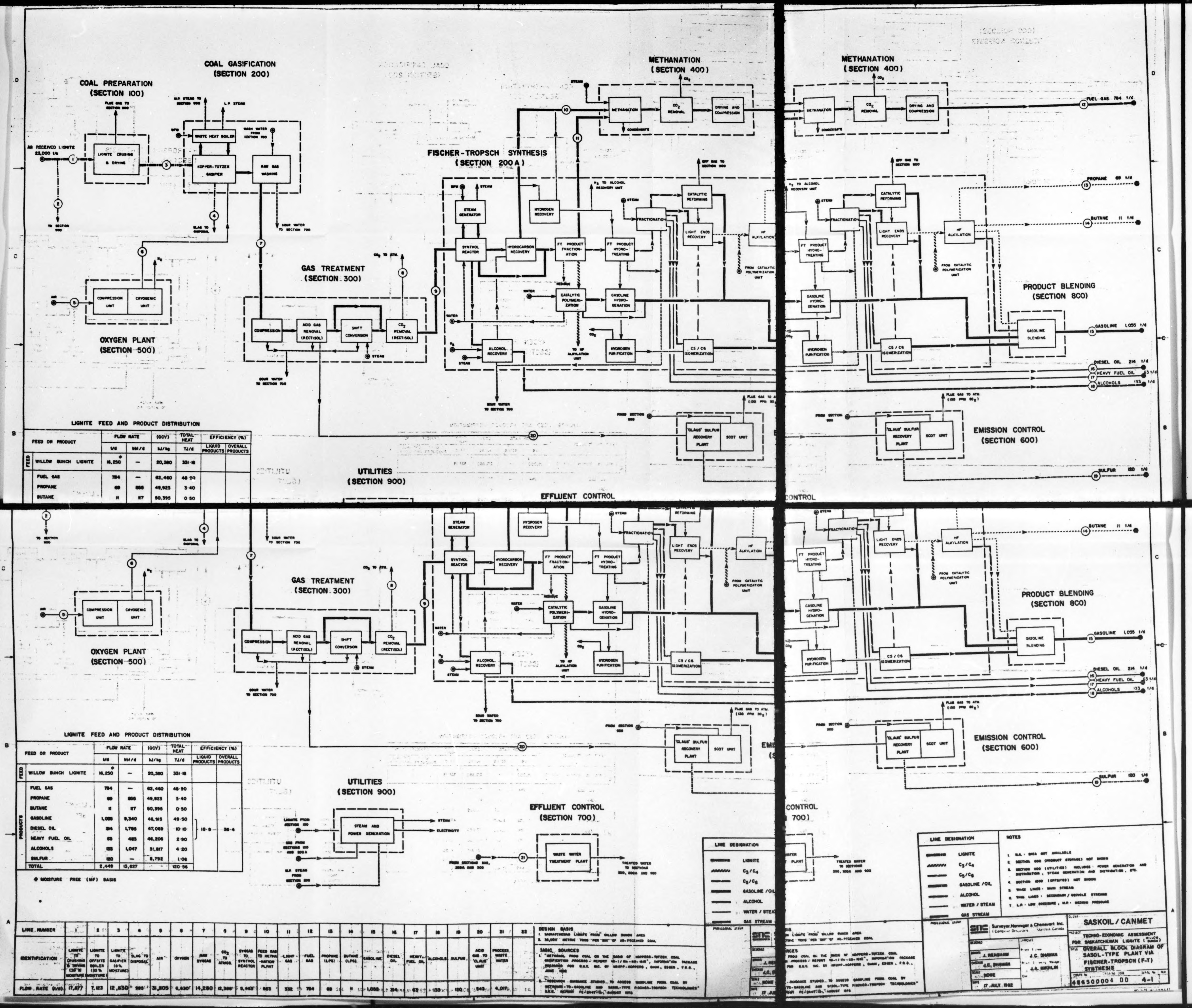
DESIGNED	
DRAWN	J. RENSHAW
CHECKED	J.C. DHAWAN
SCALE	NONE
DATE	17 MAY 1982
APPROVED	Project Engineer J.C. DHAWAN Project Engineering Manager J.A. MIKHLIN Client

CLIENT SASKOIL / CANMET

PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)

TITLE OVERALL BLOCK DIAGRAM OF COAL-BASED METHANOL-TO-GASOLINE (M-MTG) PLANT

CONTR. No.	DRAWING CODE	SERIAL No.	REV.
486500004	00	3-1	



LIGNITE FEED AND PRODUCT DISTRIBUTION

	FEED OR PRODUCT	FLOW RATE		(GCV)	TOTAL HEAT	EFFICIENCY (%)	
		t/d	bbl/d	kJ/kg	TJ/d	LIQUID PRODUCTS	OVERALL PRODUCTS
FEED	WILLOW BUNCH LIGNITE	16,250 *	—	20,380	331·18		
PRODUCTS	FUEL GAS	784	—	62,460	48·90		
	PROPANE	69	856	49,923	3·40		
	BUTANE	11	117	50,395	0·50		
	GASOLINE	1,055	9,340	46,915	49·50	18·9	36·4
	DIESEL OIL	214	1,785	47,069	10·10		
	HEAVY FUEL OIL	63	483	46,206	2·90		
	ALCOHOLS	133	1,047	31,817	4·20		
	SULFUR	120	—	8,792	1·06		
	TOTAL	2,449	13,627		120·56		

* MOISTURE FREE (MF) BASIS

LINE NUMBER	IDENTIFICATION	FLOW RATE (t/d)
1	LIGNITE TO CRUSHING & DRYING (36 % MOISTURE)	17,877
2	LIGNITE TO OFFSITE BOILER (35 % MOISTURE)	7,123
3	LIGNITE TO GASIFIER (8 % MOISTURE)	12,630
4	SLAG TO DISPOSAL	995
5	AIR	31,805
6	OXYGEN	6,830
7	RAW SYNGAS	14,280
8	CO_2 TO ATMOS.	12,386
9	SYNGAS TO SYNTHOL REACTOR	5,463
10	FEED GAS TO METHANATION PLANT	883
11	LIGHT GAS	332
12	FUEL GAS	784
13	PROPANE (LPG)	69
14	BUTANE (LPG)	11
15	GASOLINE	1,055
16	DIESEL OIL	214
17	HEAVY FUEL OIL	63
18	ALCOHOLS	133
19	SULFUR	120
20	ACID GAS TO 'CLAUS' UNIT	543
21	PROCESS WASTE WATER	4,017
22		

DESIGN BASIS
1. SASKATCHEWAN LIGNITE FROM WILLOW BUNCH AREA
2. 25,000 METRIC TONS PER DAY OF AS-RECEIVED COAL

BASIC SOURCES
1. "METHANOL FROM COAL ON THE BASIS OF KOPPERS-TOTZEK COAL GASIFICATION PROCESS - REPORT [illegible]", INFORMATION PACKAGE PREPARED FOR S.N.C. INC. BY KRUPP-KOPPERS, GmbH, ESSEN, F.R.G., JUNE [illegible]
2. "RESEARCH - GUIDANCE STUDIES TO ASSESS GASOLINE FROM COAL BY METHANOL-TO-GASOLINE AND SASOL-TYPE FISCHER-TROPSCH TECHNOLOGIES" D.O.E. REPORT FE-[illegible], AUGUST [illegible]

LINE DESIGNATION	
▬	LIGNITE
∿∿∿	C_3/C_4
▬	C_5/C_6
▬	GASOLINE / OIL
▬	ALCOHOL
—	WATER / STEAM
▬	GAS STREAM

NOTES
1. N.A. : DATA NOT AVAILABLE
2. SECTION 800 (PRODUCT STORAGE) NOT SHOWN
3. SECTION 900 (UTILITIES) INCLUDES : POWER GENERATION AND DISTRIBUTION, STEAM GENERATION AND DISTRIBUTION, ETC.
4. SECTION 1000 (OFFSITES) NOT SHOWN
5. THICK LINES : MAIN STREAM
6. THIN LINES : SECONDARY / RECYCLE STREAMS
7. L.P. : LOW PRESSURE, M.P. : MEDIUM PRESSURE

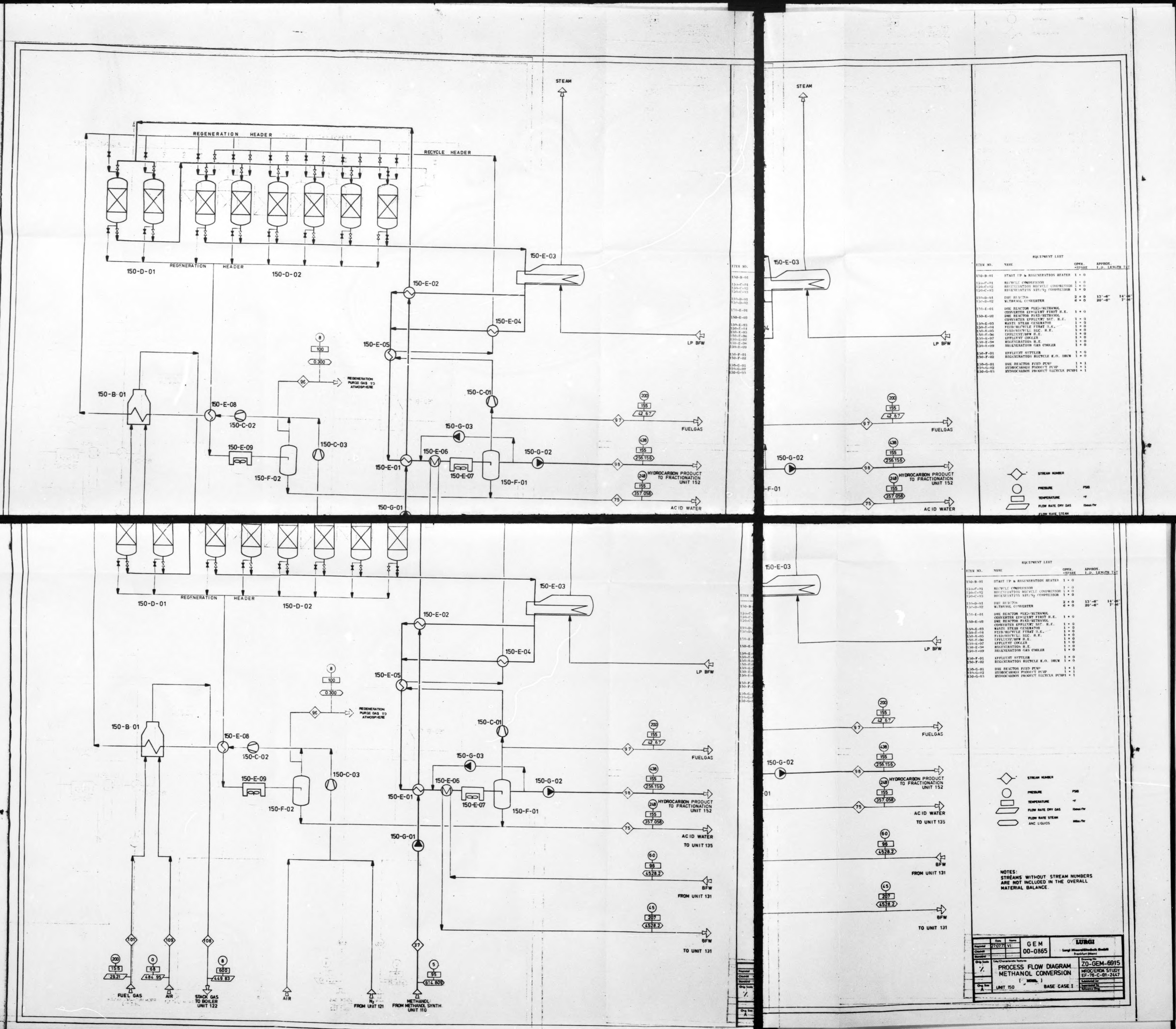
PROFESSIONAL STAMP

SNC Surveyer, Nenniger & Chenevert Inc.
I Compagnie Desjardins Montreal Canada

DESIGNED
DRAWN J. RENSHAW
CHECKED J.C. DHAWAN
SCALE NONE
DATE 17 JULY 1982

APPROVED
Project Engineer J.C. DHAWAN
Project Engineering Manager J.A. MICHLIN

CLIENT SASKOIL / CANMET
PROJECT TECHNO-ECONOMIC ASSESSMENT FOR SASKATCHEWAN LIGNITE (WILLOW BUNCH)
TITLE OVERALL BLOCK DIAGRAM OF SASOL-TYPE PLANT VIA FISCHER-TROPSCH (F-T) SYNTHESIS
486500004 00 4-1



EQUIPMENT LIST

ITEM NO.	NAME	OPER. +SPARE	APPROX. I.D.	LENGTH T-T
150-B-01	START UP & REGENERATION HEATER	1 + 0		
150-C-01	RECYCLE COMPRESSOR	1 + 0		
150-C-02	REGENERATION RECYCLE COMPRESSOR	1 + 0		
150-C-03	REGENERATION AIR/N_2 COMPRESSOR	1 + 0		
150-D-01	DME REACTOR	2 + 0	13'-6"	14'-6"
150-D-02	METHANOL CONVERTER	6 + 0	20'-6"	7'-6"
150-E-01	DME REACTOR FEED/METHANOL CONVERTER EFFLUENT FIRST H.E.	1 + 0		
150-E-02	DME REACTOR FEED/METHANOL CONVERTER EFFLUENT SEC. H.E.	1 + 0		
150-E-03	WASTE STEAM GENERATOR	1 + 0		
150-E-04	FEED/RECYCLE FIRST H.E.	1 + 0		
150-E-05	FEED/RECYCLE SEC. H.E.	1 + 0		
150-E-06	EFFLUENT/BFW H.E.	1 + 0		
150-E-07	EFFLUENT COOLER	1 + 0		
150-E-08	REGENERATION H.E.	1 + 0		
150-E-09	REGENERATION GAS COOLER	1 + 0		
150-F-01	EFFLUENT SETTLER	1 + 0		
150-F-02	REGENERATION RECYCLE K.O. DRUM	1 + 0		
150-G-01	DME REACTOR FEED PUMP	1 + 1		
150-G-02	HYDROCARBON PRODUCT PUMP	1 + 1		
150-G-03	HYDROCARBON PRODUCT RECYCLE PUMP	1 + 1		

◇	STREAM NUMBER	
○	PRESSURE	PSIG
▭	TEMPERATURE	°F
▱	FLOW RATE DRY GAS	lbmol/hr
⬭	FLOW RATE STEAM AND LIQUIDS	klbs/hr

NOTES:
STREAMS WITHOUT STREAM NUMBERS ARE NOT INCLUDED IN THE OVERALL MATERIAL BALANCE.

	Date	Name
Prepared	27.07.77	VI
Checked		
Standard		

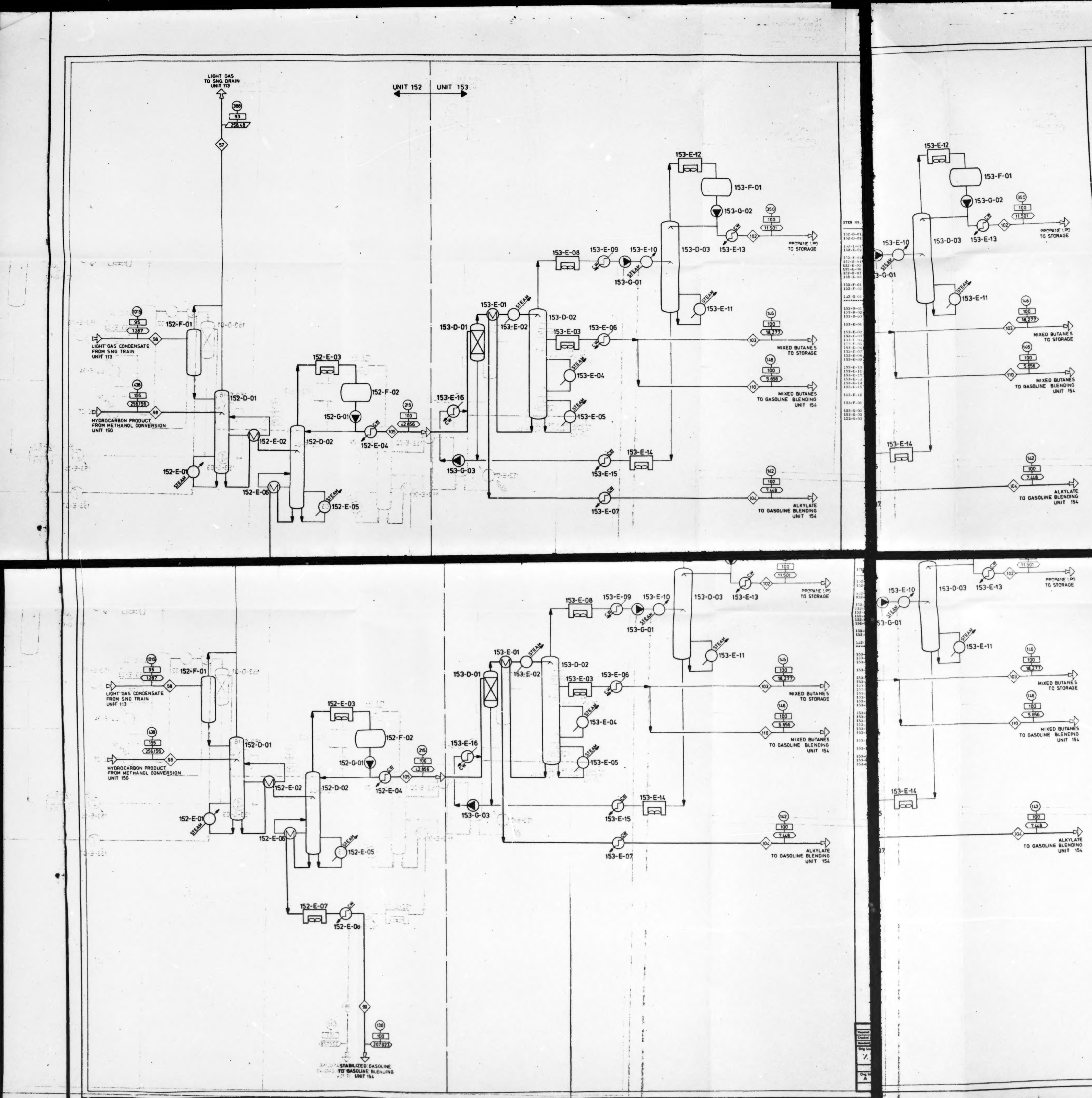
GEM
00-0865

LURGI
Lurgi Mineralöltechnik GmbH
Frankfurt (Main)

Orig. Scale ⁒

PROCESS FLOW DIAGRAM
METHANOL CONVERSION
(MOBIL)
UNIT 150
BASE CASE I

Drawing No
ZO-GEM-6915
MRDC/ERDA STUDY
EF-76-C-01-2447



EQUIPMENT LIST

ITEM NO.	NAME	OPER. + SPARE	APPROX. I.D.	LENGTH T-T
152-D-01	DEETHANIZER	1 + 0	6'-0"	80'
152-D-02	STABILIZER	1 + 0	6"-6"	105'
152-E-01	DEETHANIZER REBOILER	1 + 0		
152-E-02	DEETHANIZER SIDESTREAM/BOTTOMS H.E.	1 + 0		
152-E-03	STABILIZER OVHD CONDENSER	1 + 0		
152-E-04	STABILIZER OVHD PRODUCT COOLER	1 + 0		
152-E-05	STABILIZER BOTTOMS REBOILER	1 + 0		
152-E-06	STABILIZER SIDE REBOILER	1 + 0		
152-E-07	STABILIZER BOTTOMS COOLER	1 + 0		
152-E-08	STABILIZER BOTTOMS TRIM COOLER	1 + 0		
152-F-01	LIGHT GAS FLASH DRUM	1 + 0		
152-F-02	STABILIZER REFLUX DRUM	1 + 0		
152-G-01	STABILIZER REFLUX PUMP	1 + 1		
153-D-01	REACTOR/SETTLER	1 + 0	6'-6"	50'
153-D-02	DEISOBUTANIZER	1 + 0	9"-6"	165'
153-D-03	DEPROPANIZER	1 + 0	5'-0"	100'
153-E-01	DEISOBUTANIZER FEED/EFFLUENT H.E.	1 + 0		
153-E-02	DEISOBUTANIZER FEED HEATER	1 + 0		
153-E-03	MIXED BUTANES COOLER	1 + 0		
153-E-04	DEISOBUTANIZER SIDE REBOILER	1 + 0		
153-E-05	DEISOBUTANIZER BOTTOMS REBOILER	1 + 0		
153-E-06	MIXED BUTANES TRIM COOLER	1 + 0		
153-E-07	DEISOBUTANIZER BOTTOMS TRIM COOLER	1 + 0		
153-E-08	DEISOBUTANIZER OVHD CONDENSER	1 + 0		
153-E-09	DEISOBUTANIZER OVHD TRIM CONDENSER	1 + 0		
153-E-10	DEPROPANIZER FEED HEATER	1 + 0		
153-E-11	DEPROPANIZER REBOILER	1 + 0		
153-E-12	DEPROPANIZER OVHD CONDENSER	1 + 0		
153-E-13	DEPROPANIZER OVHD PRODUCT COOLER	1 + 0		
153-E-14	DEPROPANIZER BOTTOMS COOLER	1 + 0		
153-E-15	DEPROPANIZER BOTTOMS TRIM COOLER	1 + 0		
153-E-16	HF COOLER	1 + 0		
153-F-01	DEPROPANIZER REFLUX DRUM	1 + 0		
153-G-01	DEPROPANIZER FEED PUMP	1 + 1		
153-G-02	DEPROPANIZER REFLUX PUMP	1 + 1		
153-G-03	HF ACID RECYCLE PUMP			

◇	STREAM NUMBER	
○	PRESSURE	PSIG
▭	TEMPERATURE	°F
▱	FLOW RATE DRY GAS	Nmol/hr
⬭	FLOW RATE STEAM AND LIQUIDS	Mlbs/hr

NOTE :
STREAMS WITHOUT STREAM NUMBERS ARE NOT INCLUDED IN THE OVERALL MATERIAL BALANCE.

	Date	Name
Prepared	27777	VI
Checked		
Standard		

GEM
00-0865

LURGI
Lurgi Mineralöltechnik GmbH
Frankfurt (Main)

Drg. Scale %

PROCESS FLOW DIAGRAM
FRACTIONATION & ALKYLATION

UNITS 152/153 BASE CASE I

Drawing No. ZO-GEM-6916

MRDC/ERDA STUDY
EF-76-C-01-2447

Substitute for
Superseded by

Drg. Size A

POOR

QUALITY

ORIGINAL